Production of Ethanol from Lignocellulose

by

Simultaneous Saccharification and Fermentation

Case No. 5: Xylose Fermentation

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1 Introduction

Over the past decade, the U.S. Department of Energy's Ethanol Program, managed by the National Renewable Energy Laboratory, has used engineering analysis to help focus its research on the most promising approaches, and to develop quantitative goals for its research program. The first major analysis of enzymatic hydrolysis processes was carried out in 1985 and 1986. This study developed a "base case" process design for a separate hydrolysis and fermentation process (the best understood process in the mid-1980s), and then modified the process design to analyze a simultaneous saccharification and fermentation process (a more advanced process). Using these basic flow sheets, a series of analyses were carried out to determine the sensitivity of the ethanol production economics to the major process parameters, and to estimate the ultimate potential of each process. The results of these analyses were used to determine which enzymatic hydrolysis process would be the focus of program funded research, to estimate the potential cost of ethanol production by enzymatic hydrolysis, and to guide in-house and subcontracted research on individual processing steps. The objective of this series of reports (of which this is the fifth of the six) is to document the process designs used to make these decisions.

The 1985-1986 study, internally known as the Fuel Alcohol Technology Evaluation (FATE), first developed a base case process design for a process known as separate hydrolysis and fermentation (SHF), the enzymatic hydrolysis which was best understood at the time. In this process, lignocellulosic biomass is converted by a combination of thermochemical pretreatment and enzymatic hydrolysis to a mixed sugar stream. The six carbon sugar (glucose) component of the mixed sugars is then fermented to produce ethanol. In the more advanced simultaneous saccharification and fermentation (SSF) process, the enzymatic hydrolysis and fermentation processes are carried out in the same vessel.

The design of these processes, and the sensitivity analyses which were carried out were described in brief in a series of publications and conference papers:

- Wright, J. D., A.J. Power, and L.J. Douglas (1986). "Design and Parametric Evaluation of an Enzymatic Hydrolysis Process (Separate Hydrolysis and Fermentation)." *Biotechnology and Bioengineering Symposium Series, No. 17*. Scott, C. D., ed.
- Wright, J.D. (1987). Ethanol from Lignocellulose-An Overview. *Energy Progress*, June 1988, pp 71-78.
- Torget, R. M. Himmel, J.D. Wright, and K. Grohmann (1988). "Initial Design of a Dilute Sulfuric Acid Pretreatment Process for Aspen Wood Chips" *Applied Biochemistry and Biotechnology*, **18**, The Humana Press, C.D Scott, ed.
- Wright, J.D., C.E. Wyman, and K. Grohmann (1988). "Simultaneous Saccharification and Fermentation of Lignocellulose: Process Evaluation"

Applied Biochemistry and Biotechnology, **18**, The Humana Press, C.D Scott, ed.

- Wright, J.D. (1988). "Ethanol from Biomass," *Chemical Engineering Progress*, August 1988.
- Hinman, N., J.D. Wright, W. Hoagland, and C.E. Wyman (1989). "Xylose Fermentation," *Applied Biochemistry and Biotechnology.* **20**, pp 391-402 The Humana Press.

However, the detailed flowsheets, material and energy balances, and cost estimates which were the basis of these analyses were never published. Thus, the objective of this series of reports is to make this information easily accessible.

1.1 Overview of the Analysis Methodology

The five different enzymatic hydrolysis processes in this series of reports were all carried out on a consistent basis. The overall biomass to ethanol process flowsheet is based on a design for a dilute acid hydrolysis process prepared by Badger Engineers (Cambridge, MA) for the Solar Energy Research Institute (now NREL) in 1985 (Badger 1987). This was a detailed design whose preparation required roughly two man years of engineering effort. The basis is a plant which produces 25 million gallons per year of ethanol (except for the xylose fermentation case [Report 5], which has an annual ethanol capacity of 50 million gallons). The process design was based on the dilute sulfuric acid hydrolysis results obtained at Dartmouth University in the late 1970s and early 1980s (Kwarteng 1983). In the early 1980s, the yields and sugar concentrations produced by enzymatic and acid hydrolysis processes were similar, and therefore, an overall process design optimized for an enzymatic hydrolysis process was not greatly different than one optimized for acid hydrolysis processes.

The general methodology of the FATE study analyses was to use the overall process design taken from the Badger study, and modify only those portions necessary to carry out each process analysis. To facilitate this process, a mathematical model of the entire process was developed on a Lotus 123 Spreadsheet^{TM}. The model contained a material balance which tracked approximately 20 different chemical components through roughly 30 processing steps. From the material balance, an energy balance was calculated, and the cost of the major capital equipment was estimated. From this data, a cost of production was calculated.

The cost of the major equipment was estimated using the ICARUS^{TM} computer-aided cost estimating program, which produces a detailed time-and-materials cost estimate. For a completely defined process, the accuracy of the estimating program is $\pm 10\%$. Thus, for the SHF and SSF base cases, the limiting accuracy of the estimates is the knowledge of the process, not the cost estimating methodology. In the sensitivity studies, capital costs were calculated by multiplying the cost of the capital equipment for a given process area in the base case by the ratio of the new and base case flow rates, and applying a suitable scaling exponent (n) to the ratio:

$$\textit{Final Cost} = \textit{Base Cost} \times \left(\frac{\textit{Final Flow Rate}}{\textit{Base Case Flow Rate}}\right)^n \qquad \mathbf{(1)}$$

where the value of n reflects the type of equipment which predominates in the process area. As these calculations were carried out using a single exponent and flow rate ratio for an entire process area, the capital cost estimates in the sensitivity studies were not nearly as well known as in the base cases.

The spreadsheet was modified several times. The first modification was to replace the acid hydrolysis section with sections modeling the pretreatment, enzyme production, and enzymatic hydrolysis processes. The process schematic for the remaining sections (feedstock handling, fermentation, distillation, environmental, and offsites) remained unchanged. The resulting spreadsheet was then used to prepare the three SHF cases. The process flow diagram for these cases was essentially identical. The only differences between the cases were the yield of six carbon sugars from the hydrolysis process, and the sugar concentration in the liquid hydrolyzate stream. (Improvements in these parameters of course result in major improvements in the amount of ethanol which can be produced from a given amount of wood, large reductions in the size and cost of much of the downstream equipment, and in greatly reduced energy consumption). The SHF analysis was carried out in 1985.

The spreadsheet was modified a second time in 1987 to replace the separate enzymatic hydrolysis and fermentation sections with an integrated hydrolysis-fermentation reactor, and to replace the steam explosion pretreatment used in the previous analyses with a higher yield, lower temperature dilute acid pretreatment. By integrating the hydrolysis and fermentation processes, the sugars produced by the hydrolysis of the cellulose are immediately converted by yeast into ethanol. Thus, the sugars do not inhibit the cellulase enzymes which carry out the hydrolysis process, and the amount of enzyme needed to carry out the hydrolysis is greatly reduced.

In 1988 the spreadsheet was modified again to add a xylose fermentation process. By this time, significant progress had been made on xylose fermentation processes at NREL and other laboratories. While it was not yet clear which xylose fermentation organism would be the ultimate winner, it was clear that xylose fermentation processes would be successfully developed. Because almost one third of the sugars in hardwood and herbaceous lignocellulosic materials are five carbon sugars (of which xylose is the major component), xylose fermentation has the potential to greatly increase the process yield.

It is important to recognize that the analyses presented in this series of reports reflect the knowledge available during the mid to late 1980s. The purpose of this series of reports is to document the analyses carried out in that time period. Thus, we have not attempted to update the original flow sheets to take into account the research results of the last few years.

1.2 Overview of the Reports

An important result of the engineering analyses was a series of cases which showed how the price of ethanol production changed as a function of the technological maturity (Figure 1.1).

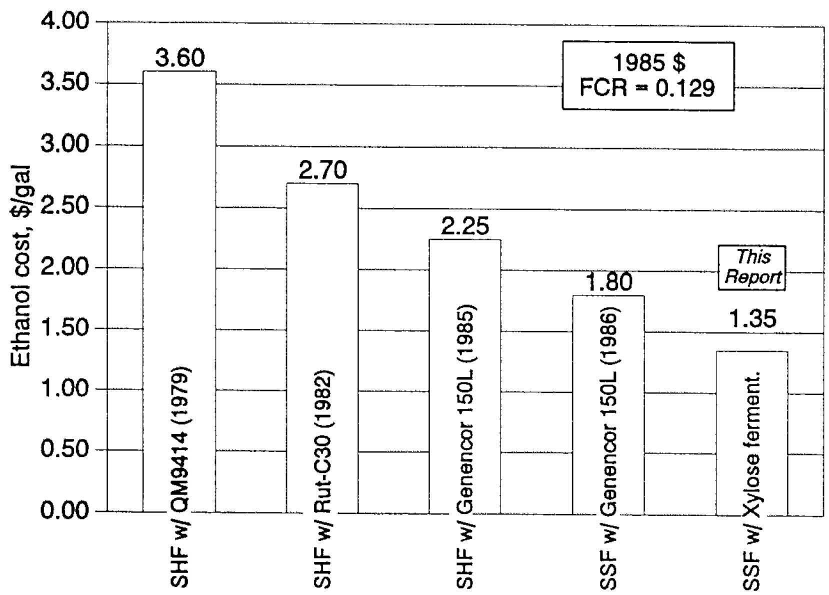


Figure 1.1 Ethanol production cost as a function of process maturity.

The first case describes the status of enzymatic hydrolysis technology in 1979, when the best available cellulase enzyme was the QM9414 mutant developed at the U.S. Army Natick Laboratories. This is a SHF process, and the projected ethanol production cost is $3.60/gallon. This case was the subject of Report 1. The second case (Report 2), was identical to the first, except that it reflected the improved hydrolysis yields and sugar concentrations achieved with the Rut-C30 cellulase enzyme developed at Rutgers in 1982. The second case is the actual "base case" enzymatic hydrolysis process. Thus, all of the other cases are essentially sensitivity studies using process flow sheets and economics derived from this case. The estimated cost of production for this case is $2.70/gallon. The third case (Report 3) uses the hydrolysis parameters achieved with the Genencor 150L cellulase mutant. The estimated cost of ethanol production is $2.25/gallon. The design of these processes is summarized by Wright, Power and Douglas (1986).

The fourth case (Report 4) is the initial SSF case, and reflects the performance achieved with the Genencor 150L enzyme at the Solar Energy Research Institute in 1985. The estimated cost of production is $1.80/gallon. This was the first SSF case analyzed (Wright, Wyman, and Grohmann 1988).

This case, the fifth case (Report 5) is identical to the previous case, except that the xylose sugars produced during the pretreatment process are fermented to ethanol (increasing the overall conversion of biomass to ethanol by roughly 50%) instead of being concentrated and burned in the boiler (Hinman, Wright, Hoagland and Wyman 1989). Additional processing steps are introduced into the dilute acid pretreatment to properly prepare the feedstock for xylose fermentation.

The sixth report in this series is an executive summary, which will describe the evolution of the processes analyzed between 1984 to 1988 at NREL.

While the reports in this series do not cover acid hydrolysis processes, analyses of both enzyme- and acid-based lignocellulose to ethanol processes were carried out at the same time. The flow sheets and cost estimates for the acid processes are consistent with those presented in this series of reports. The acid hydrolysis analyses are described in:

- Wright, J. D. and A.J. Power (1986). "Comparative Technical Evaluation of Acid Hydrolysis Processes for Conversion of Cellulose to Alcohol." Proceedings of the Conference on *Energy from Biomass and Wastes X*, Institute of Gas Technology, 7-10 April 1986, Washington, DC.
- Wright, J. D. and A. J. Power (1986). "Concentrated Halogen Acid Hydrolysis Processes for Alcohol Fuel Production." *Biotechnology and Bioengineering Symposium Series, No. 15*, 511-532. Charles D. Scott, ed. New York: John Wiley & Sons.

1.3 Organization of the Report

Following this introduction, Section 2 presents an overview of the SSF process. In this section, we first describe the chemical composition of lignocellulose and how the feedstock's chemical structure affects the conversion process. We then present an overview of the entire process. Section 3 presents a step by step description of the process, including all the major process parameters and equipment. Finally, Section 4 presents the capital and operating costs, and the overall process economics.

2 Overall Process Design

A block flow diagram of the overall cellulose to ethanol process is shown in Figure 2.1. Note that only the pretreatment, SSF, xylose fermentation, and enzyme production blocks are specific to enzymatic hydrolysis; the rest of the process is essentially identical to the acid hydrolysis process described by Badger (1984).

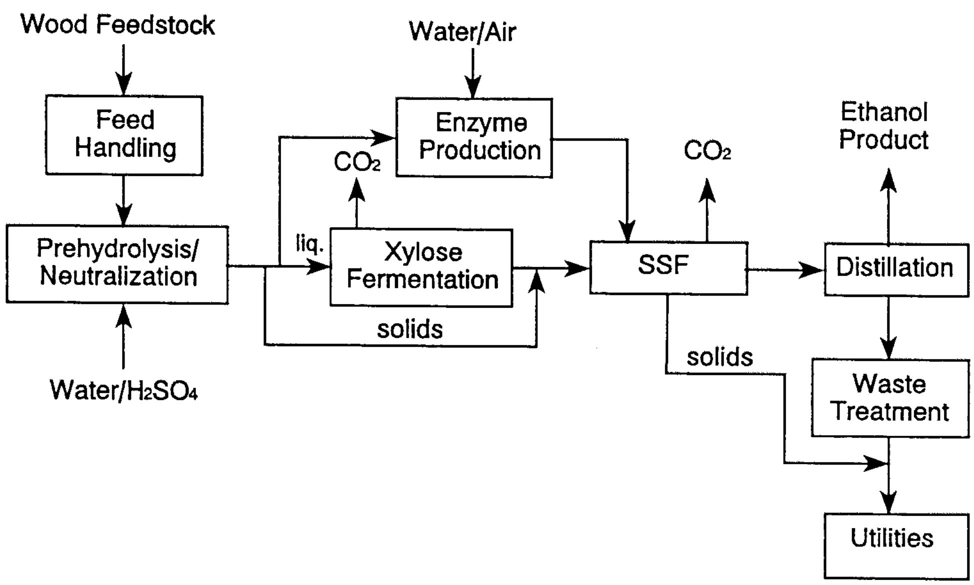


Figure 2.1 SSF overall block flow diagram

After the feedstock is received at the facility, it is pretreated to render the cellulose digestible to the hydrolytic enzymes; pretreatment also serves to fractionate the feedstock into its three major components (cellulose, hemicellulose, and lignin). The xylose portion of the sugars and acids solubilized during pretreatment is fermented to ethanol; the resulting stream is then combined with the solids and liquids entering the SSF reactor. The enzymes are produced, using a small fraction of the pretreated feed as a carbon source, and the resulting enzyme sent to the SSF step. In the SSF step, the polymeric cellulose is converted to glucose, which is then fermented to ethanol. The ethanol is then purified by distillation. The organic acids, and other soluble products from the preceding steps are concentrated by multi-effect evaporation, mixed with the lignin, and burned to produce steam and electric power to run the facility.

2.1 Feedstock Characteristics

The major fractions of lignocellulose are cellulose, hemicellulose, and lignin. Crystalline cellulose, amorphous cellulose, and hemicellulose (the carbohydrate portions of lignocellulose) can be hydrolyzed to their component sugars. Cellulose is a linear polymer of D-glucose (a six carbon sugar) molecules held together by beta-glycosidic bonds. Cellulose fibers are arranged in bundles of parallel chains in which adjacent chains are bound together by hydrogen bonding between hydroxyl groups and hydrogen atoms, forming a crystalline material with great mechanical strength and high chemical stability. Because the bond between the glucose units is the weak link in the chain, the polymer can be broken (hydrolyzed) into its component sugars (Figure 2.2).

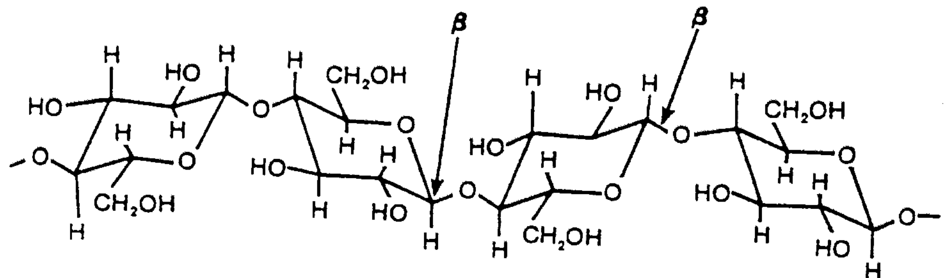


Figure 2.2 Structure of crystalline cellulose.

Hemicellulose is a polymer consisting primarily of five carbon sugars (pentosans or xylans), six carbon sugars, and organic acids (Figure 2.3). The six carbon sugars are readily fermentable to ethanol, but standard industrial yeasts can not ferment the five carbon sugars, until the use of a mixed culture and a separate xylose fermentation step in this case (Report 5). Unlike cellulose, the structure and composition of hemicellulose may vary widely between species. Hemicellulose is not crystalline and is readily hydrolyzed.

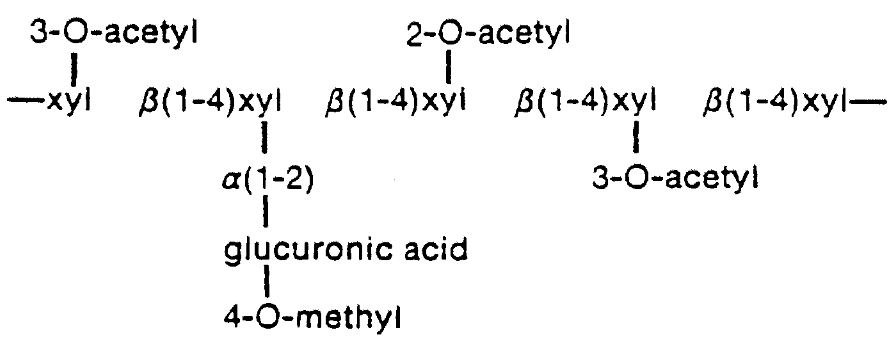


Figure 2.3 Structure of the xylan fraction of hardwood hemicellulose

Lignin, the non-carbohydrate portion of the cell wall, is chemically bonded to and mixed with the hemicellulose. Lignin is a phenolic polymer (Figure 2.4 and Figure 2.5), and can not be converted to fermentable sugars (Wenzl 1970).

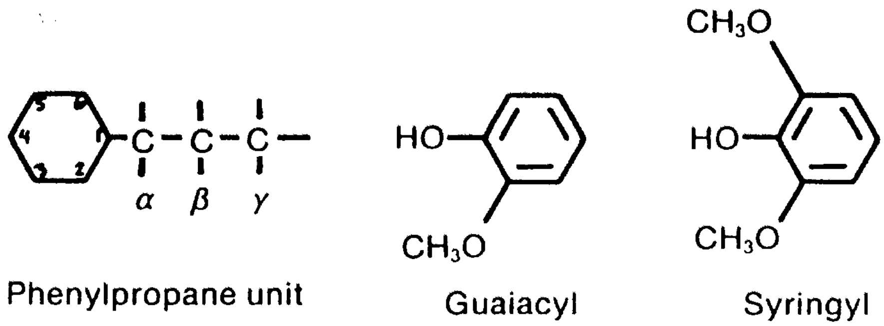


Figure 2.4 Monomer units in lignin.

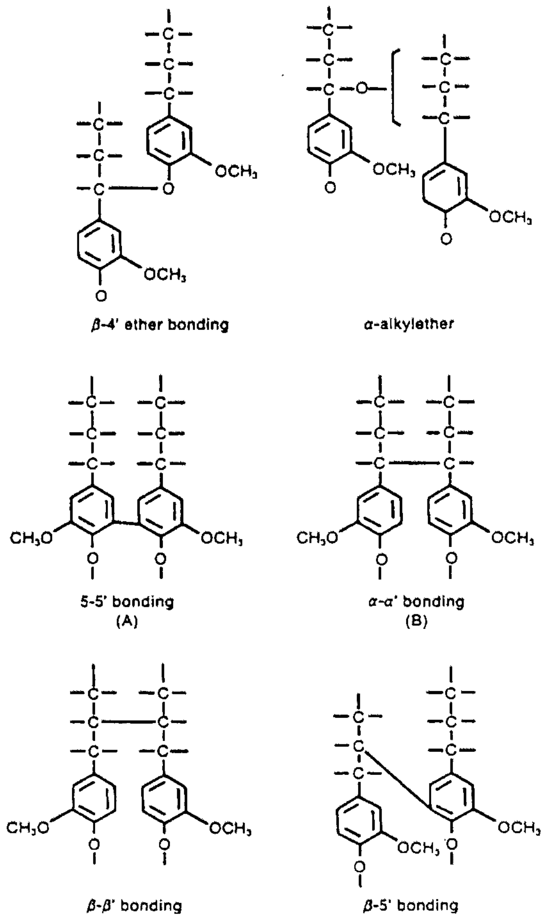


Figure 2.5 Ether and carbon-carbon bonds in lignin.

2.2 Overall Process Description

Before delving into the details of enzymatic hydrolysis, it is useful to describe the overall process. A more detailed flow sheet, along with a simplified material balance is shown in Figure 2.6.

Fresh mixed hardwood (primarily oak) wood chips (Table 2.1) are delivered by rail or truck, washed, and transferred to the wood pile. Chips are stored for a maximum of two weeks and then sent to the pretreatment section. Here, chips are subjected to a dilute (1.1%) sulfuric acid solution at 160°C. The reactor has no free liquid phase and reduces sugar dilution and energy consumption by minimizing the amount of water processed. At these conditions, 93% of the xylan is hydrolyzed, resulting in a fully digestible cellulose pulp (Torget *et al.* 1988). The pretreated solids are then neutralized with lime and sent to the SSF section; approximately 2.7% of the solids are separated and used as a substrate for enzyme production. Pretreatment also produces a liquid stream containing approximately 6% xylose which is then delivered to the xylose fermentation section. The xylose is fermented to ethanol at 70% of the theoretical efficiency, and the resulting stream is fed to the SSF reactor.

Table 2.1 Wood composition used in the analysis.

Component	Weight (%, dry basis)	Gross heat of combustion (kJ/kg)[1]	Fraction of total energy content (%)
Cellulose	46.2%	17,350	44.7%
Hemicellulose	24.0%	16,676	22.3%
Lignin	24.0%	24,702	33.0%
Other	5.8%	0	0.0%
Total		17,946	

[1] Domalski and Milne (1987)

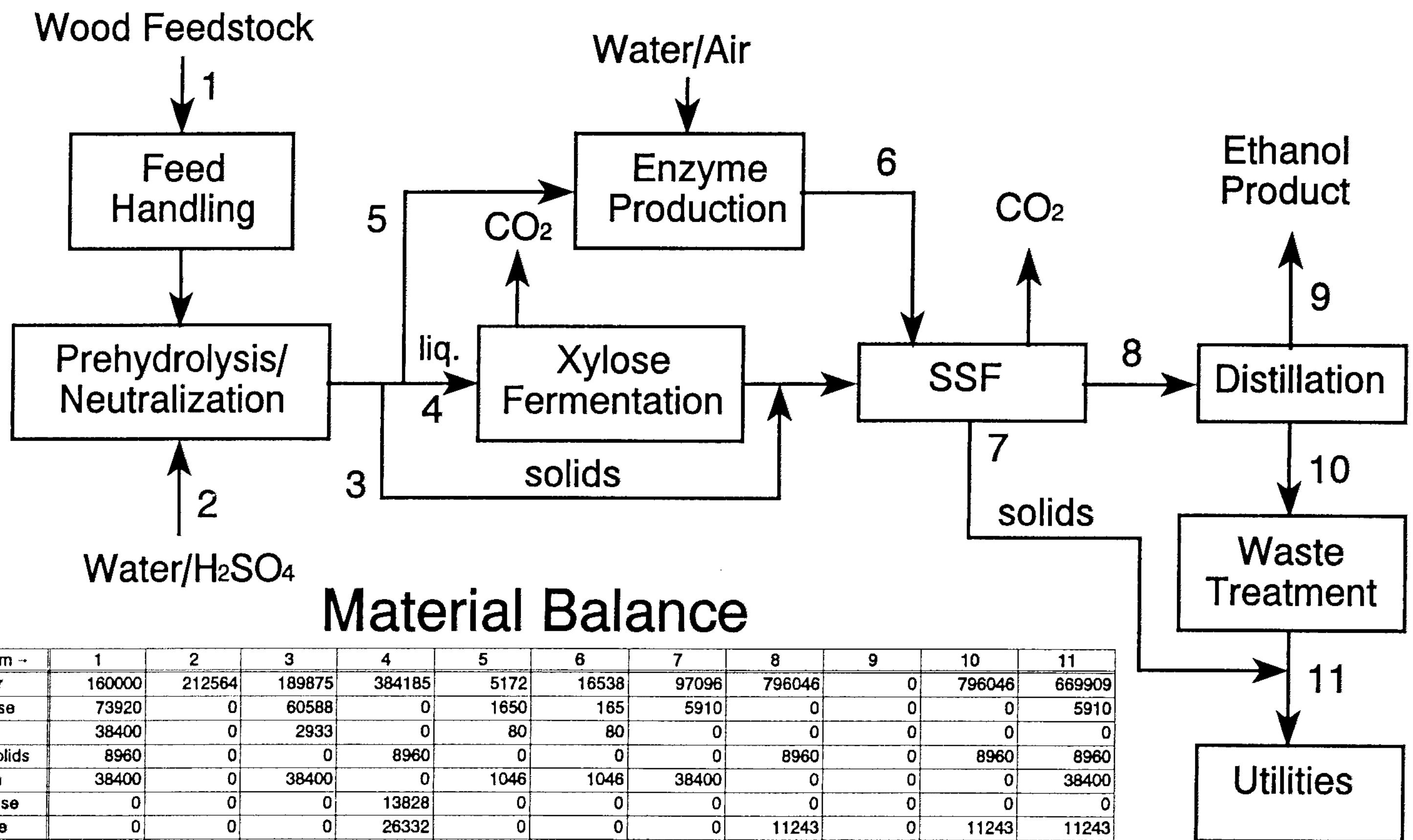
The cellulase enzyme is produced by the fungal mutant Genencor 150L using a fed batch production system design developed at the University of California, Berkeley (Wilke and Blanch 1985). The production is an aerobic process which uses pretreated wood as the primary carbon source. Productivity is approximately 83 IU/l-hr, with a final cellulase titre of 24 IU/ml and a residence time of just over 10 days.

The pretreated, neutralized wood is charged directly to SSF, which is carried out in large, agitated, carbon steel batch reactors (Wright *et al.* 1988). Whole enzyme broth from the enzyme production fermenters (including cellulose not digested by the fungus during enzyme production) is added to the solids, and the system diluted to 10% cellulose by weight. The residence time is seven days, with an enzyme loading of 7 IU/g cellulose. Of the total cellulose feed, 88% is hydrolyzed. Of this, 90% is fermented to ethanol and 10% is converted to cell mass and byproducts. The final ethanol concentration is 4.8%. All of the hemicellulose which was not solubilized in the pretreatment step is hydrolyzed to simple sugars (*e.g.* xylose).

Dilute beer from SSF and xylose fermentation is concentrated to 40 weight percent in the beer column. The flow of stillage (unconverted xylose, waste organics) from the bottom of the column is pumped to evaporation. Distillate is fed to the rectification column, which concentrates the ethanol to its azeotropic composition. The bottoms stream is recycled to the beer column. Several components, collectively termed fusel oils, tend to accumulate within the rectification column and are removed in a sidedraw purge. Azeotropic ethanol is dehydrated to 199+ proof by molecular sieve absorption.

In the evaporation section, a three-stage evaporator is used to concentrate stillage to produce a solution of 60% mixed organics in water, which is burned as a liquid fuel to produce steam. About half of the recovered condensate is recycled to the process and about half is sent to anaerobic digestion.

Two boilers are designed to burn solid, liquid, and gaseous fuel byproducts from the process units in order to generate 1100 psia steam for use in the process and for electrical power generation. Gas and liquid fuels are burned directly but lignin and undigested cellulose from hydrolysis is fed to a Flakt type drying system which dries and fluidizes the solids into the burners using boiler flue gas. A turbogenerator package unit is rated to take 1100 psia, 825°F steam and generate sufficient electricity to meet all the internal needs of the facility, with additional power available to export as an additional revenue stream.



Material Balance

Stream →	1	2	3	4	5	6	7	8	9	10	11
Water	160000	212564	189875	384185	5172	16538	97096	796046	0	796046	669909
Hexose	73920	0	60588	0	1650	165	5910	0	0	0	5910
Xylan	38400	0	2933	0	80	80	0	0	0	0	0
Sol solids	8960	0	0	8960	0	0	0	8960	0	8960	8960
Lignin	38400	0	38400	0	1046	1046	38400	0	0	0	38400
Glucose	0	0	0	13828	0	0	0	0	0	0	0
Xylose	0	0	0	26332	0	0	0	11243	0	11243	11243
Ethanol	0	0	0	0	0	0	0	42848	42634	214	214
H2SO4	0	3263	0	0	0	0	0	0	0	0	0
Cell Mass	0	0	0	0	0	201	3918	0	0	0	3918
Enzyme	0	0	0	0	0	655	0	655	0	655	655
Total	320000	215827	292116	433986	11835	17193	145644	860434	42634	817800	749242

Figure 2.6 Flow sheet and material balance for the separate hydrolysis and fermentation process.

3 Individual Process Step Descriptions

This section provides a process description for each major step in the overall cellulose-to-ethanol facility. Nominal design capacity is 52 million gallons per year of absolute ethanol; the feed rate (dry basis) for this case is 1920 ton/d. This is the feed rate used (and maintained constant) in the original SSF/xylose fermentation analysis, while in the other reports the feed rate was varied to produce a fixed 25 million gallons of ethanol per year. As discussed above, design of each the individual process areas is based on the Badger (1987) report; exceptions are noted. The reader is referred to the process flowsheets as appropriate for each step. At the lower right-hand corner of each flowsheet is the "title" of that drawing, to which all interconnecting streams refer.

The detailed equipment lists were developed along with the process designs, by Badger, Stone & Webster, and Chem Systems. To reconcile differences in the numbering systems used by these three, we have adopted a simplified equipment nomenclature (Table 3.1). Appendix B presents the equipment list for the entire facility, listed by process section. The column labeled "P&ID No." in Appendix B corresponds to the equipment identification number shown on the appropriate process flow diagram. For each item, the quantity, electrical requirements, critical design parameters, and purchase costs are listed. Note: the "design data" listed are for the case where hydrolysis yield is 72%, and should be multiplied by the "scale factors" shown to estimate the conditions for the hydrolysis and fermentation yields used in this report.

Table 3.1 Equipment nomenclature

Letter	Equipment
B	Boilers
C	Centrifuges
D	Dryers
F	Filters
G	General (Scales, Specialty items, *etc.*)
H	Heat Exchangers
J	Conveyors
M	Motors
P	Pumps, Compressors
R	Reactors, Towers, Columns
T	Tanks
U	Package Units

3.1 Feedstock Handling (Section 100)

The lignocellulosic feedstock is received from truck trailers (24 ton capacity) or train cars (60 ton capacity) on a continuing basis. Sufficient storage for two weeks operation is provided in the wood chip pile (Figure 3.1). Chips are off-loaded from trailers or trains into the washing flume, allowing separation of rocks and large metal items. From the flume the chips are discharged to the chip pile by the bucket elevator (J-101). The pile is managed on a first in, first out basis by one of two front end loaders (G-104A,B), the other of which feeds chips into the vibratory screen (F-103). A belt conveyor (J-102) receives chips from the vibratory screen and delivers

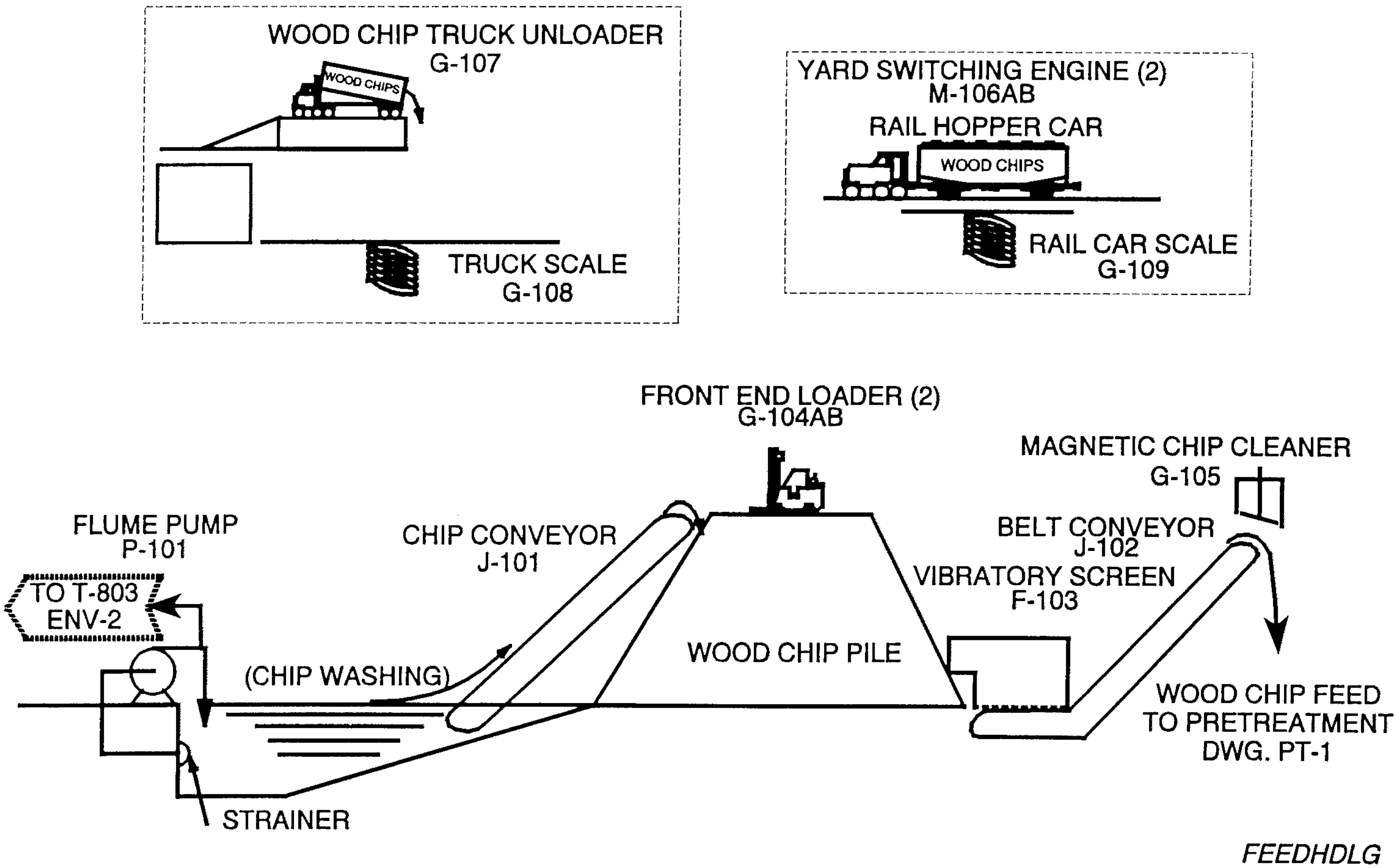


Figure 3.1 Feedstock handling

them to pretreatment, with small metal being removed by the magnetic chip cleaner (G-105).

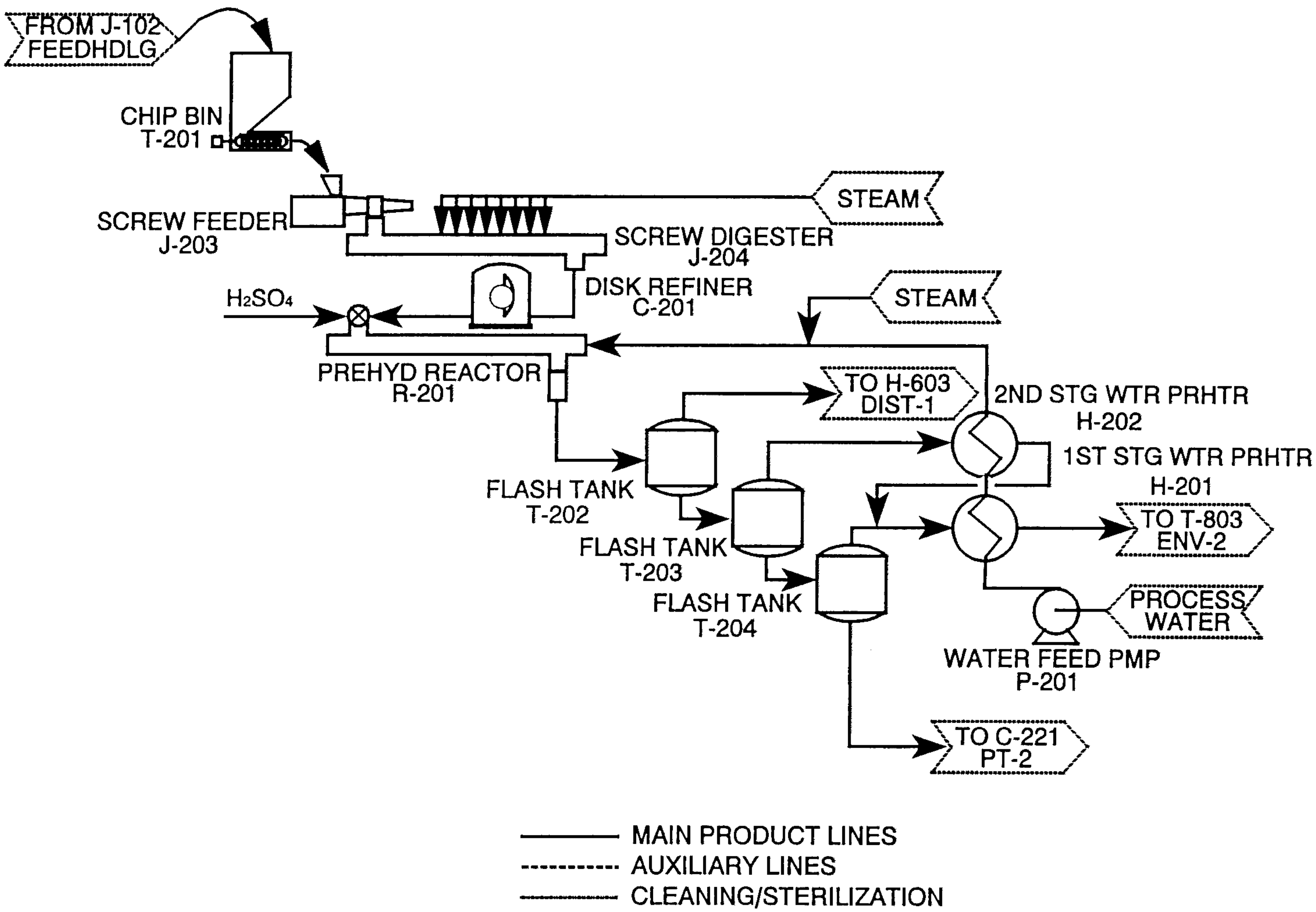
3.2 Pretreatment (Section 200)

The complete dilute acid pretreatment is shown in Figure 3.2 and Figure 3.3. The overall design of this process section is based on work by Torget (1988) with modifications by Chem Systems (1987). Mixed hardwood chips are transferred from feedstock handling to chip bin T-201, then fed from the chip bin to digester J-204 where they are mixed with 17 atm steam. The digester operates at 160°C with a residence time of ten minutes. This softens the wood and provides a pressure head for further processing. The softened wood chips are discharged into single disk refiner C-201 which reduces the particles to nominal 1 millimeter diameter short fibers. After size reduction, the wood particles are discharged through a ball valve into the prehydrolysis reactor.

The prehydrolysis reactor, R-201, is designed for a ten minute residence time at 160°C. The wood particles are mixed with water and steam to achieve a 15 percent solids concentration and then sulfuric acid is added to a concentration of 1.1 weight percent based on the water content. The water feed is preheated by exchange with flash vapors in first stage water preheater H-201 and in second stage water preheater H-202 to reach a temperature of 105°C prior to injection into the prehydrolysis reactor. All remaining heat is supplied by saturated 17 atm steam. Within the prehydrolysis reactor most of the hemicellulose fraction (92.7%) and the amorphous cellulose are converted to their respective sugars and some degradation products such as furfural. The reactor product is flashed in three stages to remove some water/furfural vapor and simultaneously recover heat from the prehydrolysis reactor.

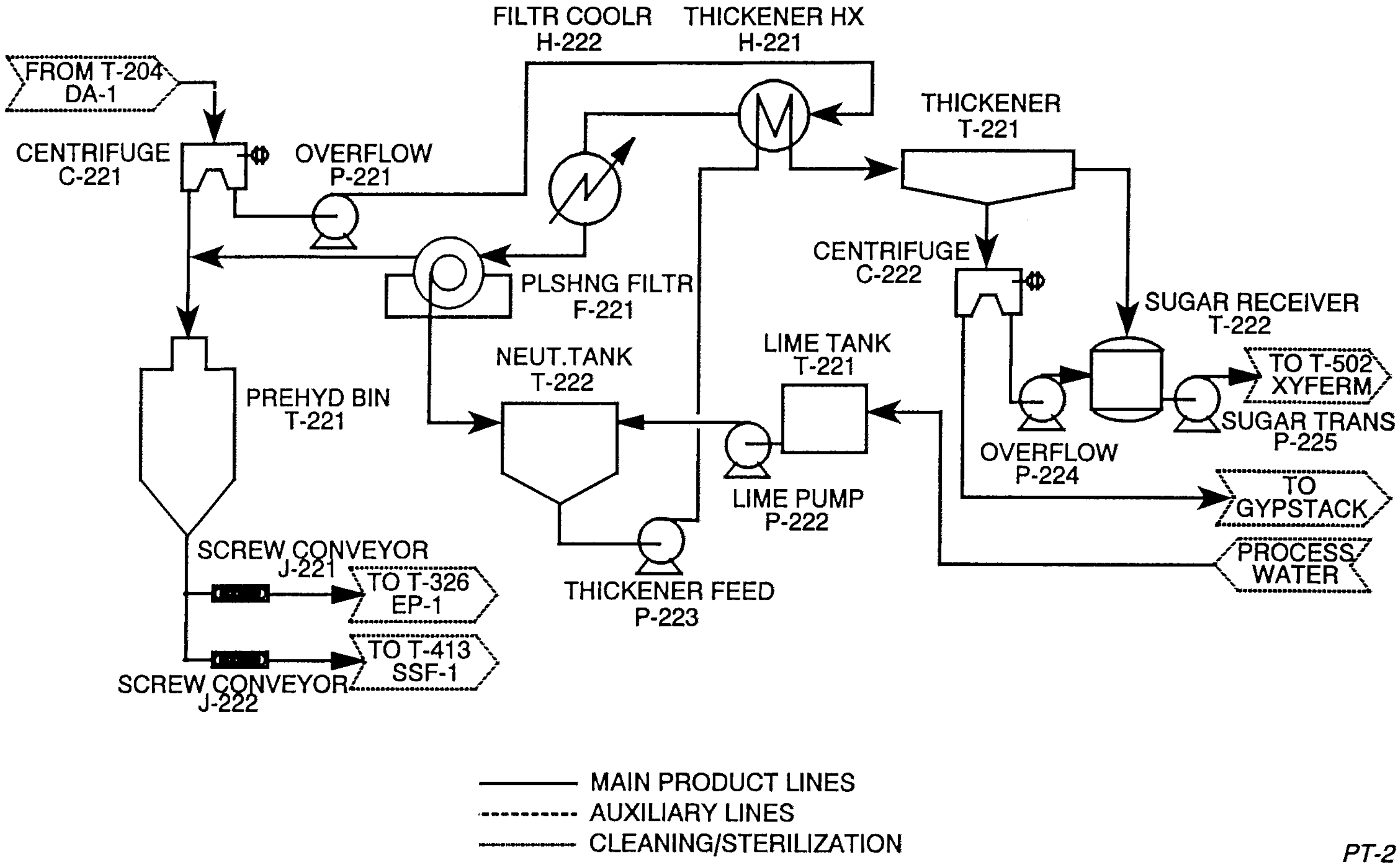
The first flash vessel, T-202, operates at 148°C. Liquid/ solid discharge proceeds to the second flash vessel, T-203. Discharge from this flash is further let down to 1.1 atm in T-204. The effluent from the third flash containing approximately 14 weight percent solids, passes to sugar separation centrifuge C-221. Flash vapors from T-202 are sent to ethanol purification in Section 600 as beer still preheat. Flash vapors from T-203 are condensed against prehydrolysis feed water in second stage water preheater H-202. This condensate is let down and combined with vapors from the third stage flash. This combined furfural/water condensate is condensed against prehydrolysis feed water in first stage water preheater H-201, and proceeds to waste treatment in Section 800.

Prehydrolyzed wood slurry is fed to centrifuge C-221 to be separated into solid and liquid streams (Figure 3.3). The overflow, which contains most of the solubles and approximately 5 percent of the solids is pumped to rotary drum polishing filter F-221 via thickener exchanger H-221 and filter cooler H-222, which cool the stream to 60°C. The bottoms from C-221, containing 35 weight percent solids, are accumulated in prehydrolyzate bin T-221. In the



PT-1

Figure 3.2 Dilute acid pretreatment, part 1



PT-2

Figure 3.3 Dilute acid pretreatment, part 2 polishing filter, approximately 95 percent of the solids are recovered and sent to T-221. The clean sugar solution proceeds to neutralization. Prehydrolyzate solids are split with 2.7 percent conveyed to enzyme production in Section 300 and the remainder conveyed to SSF. Overall the level of complexity of the pretreatment is much greater in this case than in case 4, because the solid and liquid product streams must be separated prior to xylose fermentation.

In neutralization tank T-222, the acid in the sugar solution is neutralized with calcium hydroxide solution from lime tank T-221. The neutralized sugar solution containing insoluble calcium sulfate is pumped to a thickener via thickener exchanger H-221 which reheats the stream to 83°C. In thickener T-221, the calcium sulfate solids are concentrated to 20 weight percent and fed to a solid bowl centrifuge. The clear thickener overflow is accumulated in sugar receiver T-222. Calcium sulfate centrifuge C-222 produces a high solids content salt which is dumped into a slurry transfer line and transported to the gypstack. Drainage water from the gypstack is accumulated and recirculated through the slurry transfer line to transport the calcium sulfate solids. Overflow from C-222 is pumped to sugar receiver T-222. The accumulated sugar solution from T-222 is pumped to SSF in Section 500.

3.3 Xylose Fermentation (Section 500)

The equipment required for this section is a simplified version of that required in the (glucose) fermentation sections of the more conventional separate hydrolysis and fermentation (SHF) process, as discussed in reports 1 through 3. The scale is much smaller, since only about 18 million of the 50 million total gallons of ethanol per year is derived from xylose fermentation. Equipment numbers have been left constant where possible, to facilitate comparisons between this process and the earlier reports.

The xylose fermentation section is shown in Figure 3.4. The liquid stream from pretreatment is delivered to the 14 xylose fermenters T-502, which are equipped with heat exchangers and recycle pumps. Yeast is delivered from the yeast mix tank (T-550). The ethanol product stream is fed to the SSF fermenters (T-413, SSF section) by P-554.

3.4 Enzyme Production (Section 300)

The total hydrolysis enzyme requirement is provided in the enzyme production section. The system design employs a fed batch enzyme fermentation design developed at the University of California, Berkeley (Orichowskyj 1982). The organism is a highly mutated *T. reesei*. 10-day residence time is required to achieve a final titre of 24 filter paper units/ml (FPU/ml). Fermenters are batch-fed to achieve a total cellulose concentration equivalent to a 15 wt % (150 g/1) cellulose batch operation. The fed batch system differs from a conventional batch system in that fresh solid feed is added to the enzyme production tanks periodically over the course of the total batch, resulting in a much higher total solids concentration being

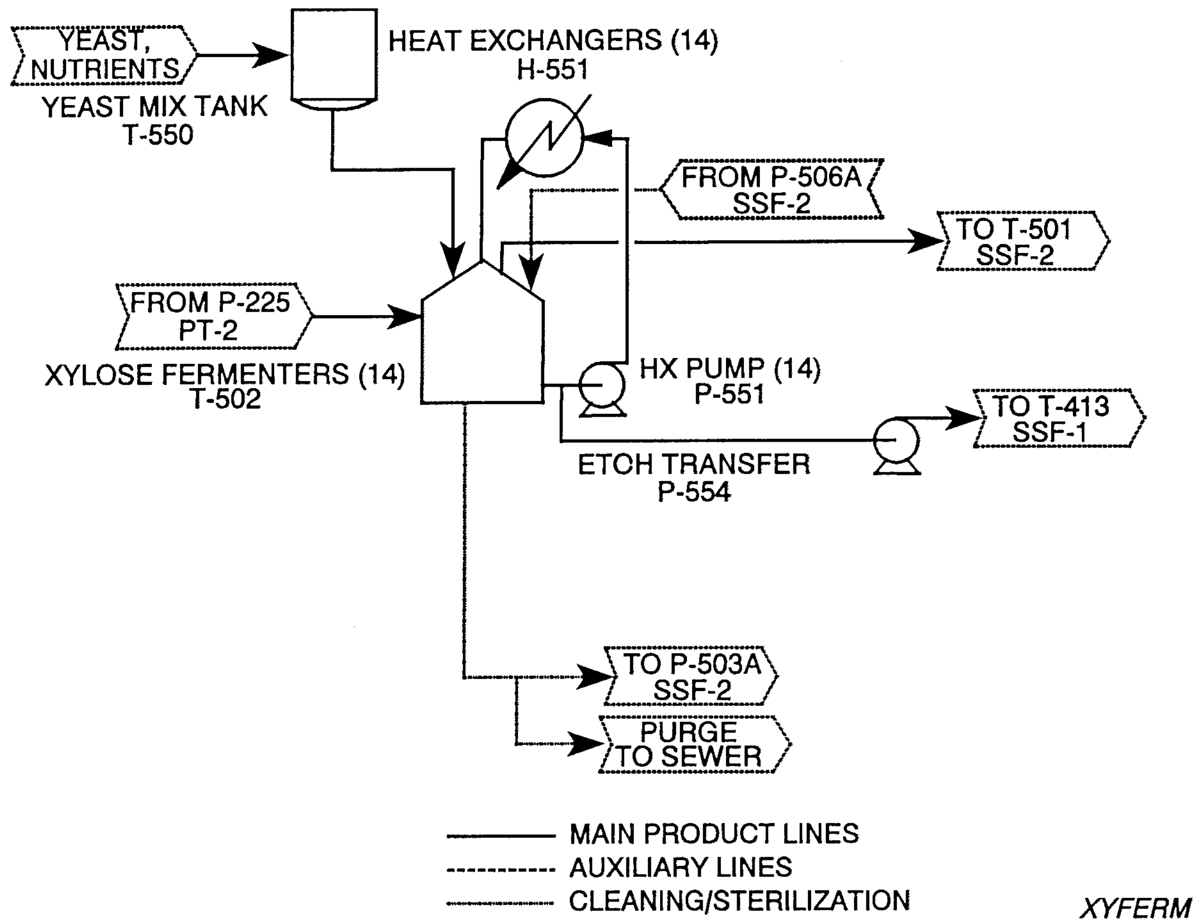


Figure 3.4 Xylose fermentation

processed. The enzyme performance is that of Genencor 150L.

The enzyme production section is shown in Figure 3.5. The initial charge of pretreated lignocellulose is conveyed into the first stage seed tank (T-326), where it is diluted with sterile water, adjusted for pH, fed nutrient, and inoculated with organism. The primary nutrient is corn steep liquor (15 g/l at 54-percent solids). Ammonia, a supplemental nitrogen source and pH regulator is added in the amount of 20 g/l (sufficient to maintain nitrogen balance for enzyme production and cell synthesis). There are three stages of seed tanks (also called inoculum fermenters), which combine to produce a small amount of enzyme to "seed" the entire set of eleven batch-fed enzyme production fermenters (T-314). These are 18-ft diameter by 60-ft-high, cone bottom, vertically stirred vessels with supplemental air sparging. The sterile air is supplied by air compressors (P-321A,B), which are driven by back-pressure steam turbines. The mixing and air sparge rates were optimized using standard general mass transfer techniques. The reaction heat is removed from the fermenter vessels by external circulation heat exchangers (H-300) cooled with chilled water. The total solids concentration in the fermenters, including recycle solids, is about 16 percent. Final enzyme concentration is 24 IU/ml which translates to an enzyme productivity of about 83 IU/liter/hour.

3.5 Simultaneous Saccharification and Fermentation (SSF) (Section 400)

The simultaneous saccharification and fermentation (SSF) section is based on the work of Wright, Wyman and Grohmann (1988). For this report, equipment from the enzyme hydrolysis (Section 400) and fermentation (Section 500) sections of the Badger report have been combined. Equipment numbers have wherever possible been left unchanged, to facilitate comparisons between this and previous reports. Clearly, the ability to accomplish both enzyme hydrolysis and ethanol production in a single vessel results in substantial reduction of hardware requirements. The magnitude will become more apparent in Section 4 of this report.

The entire SSF section is shown in Figure 3.6 and Figure 3.7. A total of eighteen fermenters (T-413) are required. In addition there are two surge bins (T-405A,B). The fermenters and surge tanks are standard pulp and paper high density stock tanks with bottom mixing. Agitation requirements are based on requirements for moving a thick pulp and paper stock.

While the hydrolysis reactors are filling, beer from xylose fermentation, makeup evaporator water, and fresh enzymes are added. The surge tanks are then prepared for the next cycle. The yield of glucose from cellulose from the hydrolysis portion of SSF is 90% (mole basis). Each fermenter has its own heat exchanger (H-501) and circulation pump (P-501). The fermenters are designed for a liquid loading of 80% of maximum capacity. The fermenters are filled on a three hour cycle. The fermentation yield is 95% of theoretical.

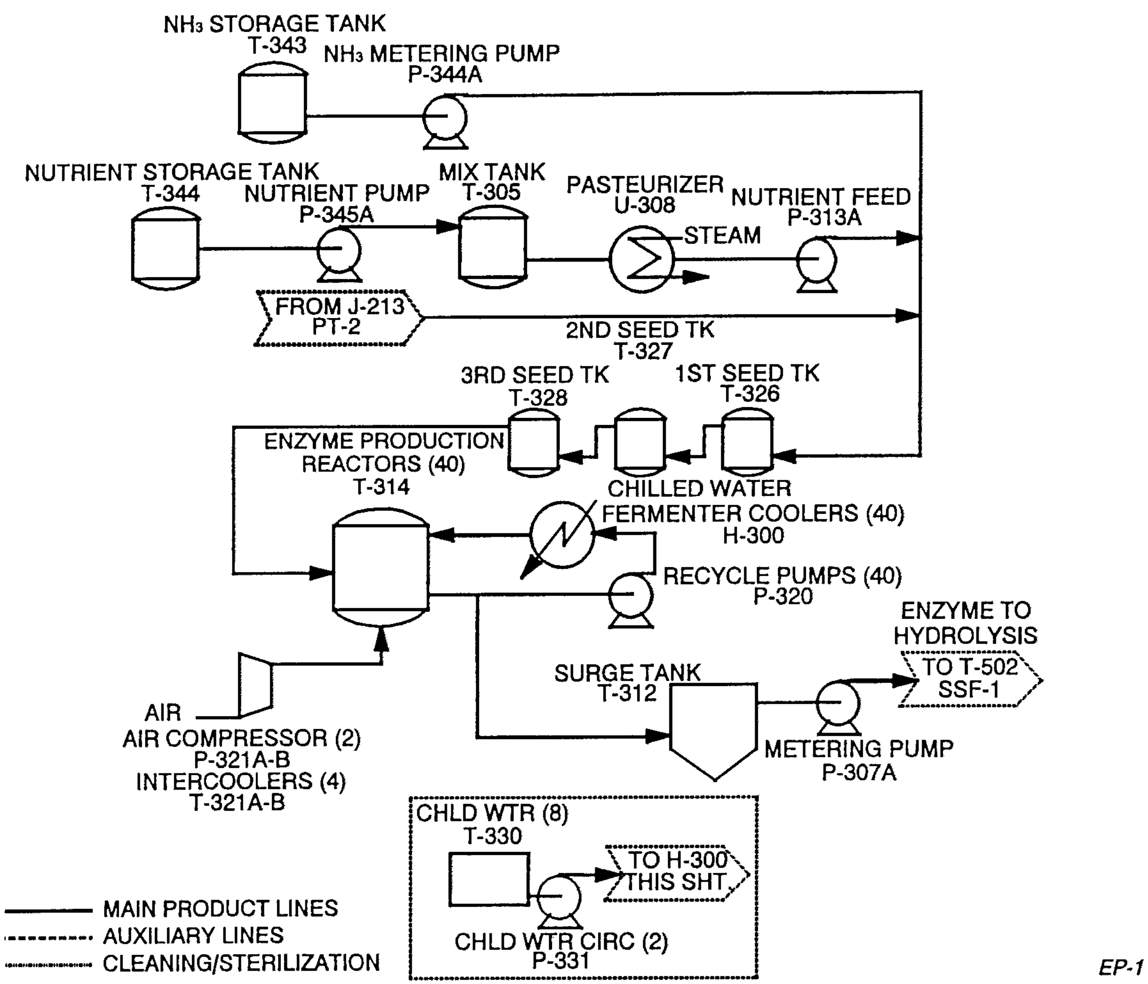


Figure 3.5 Enzyme production

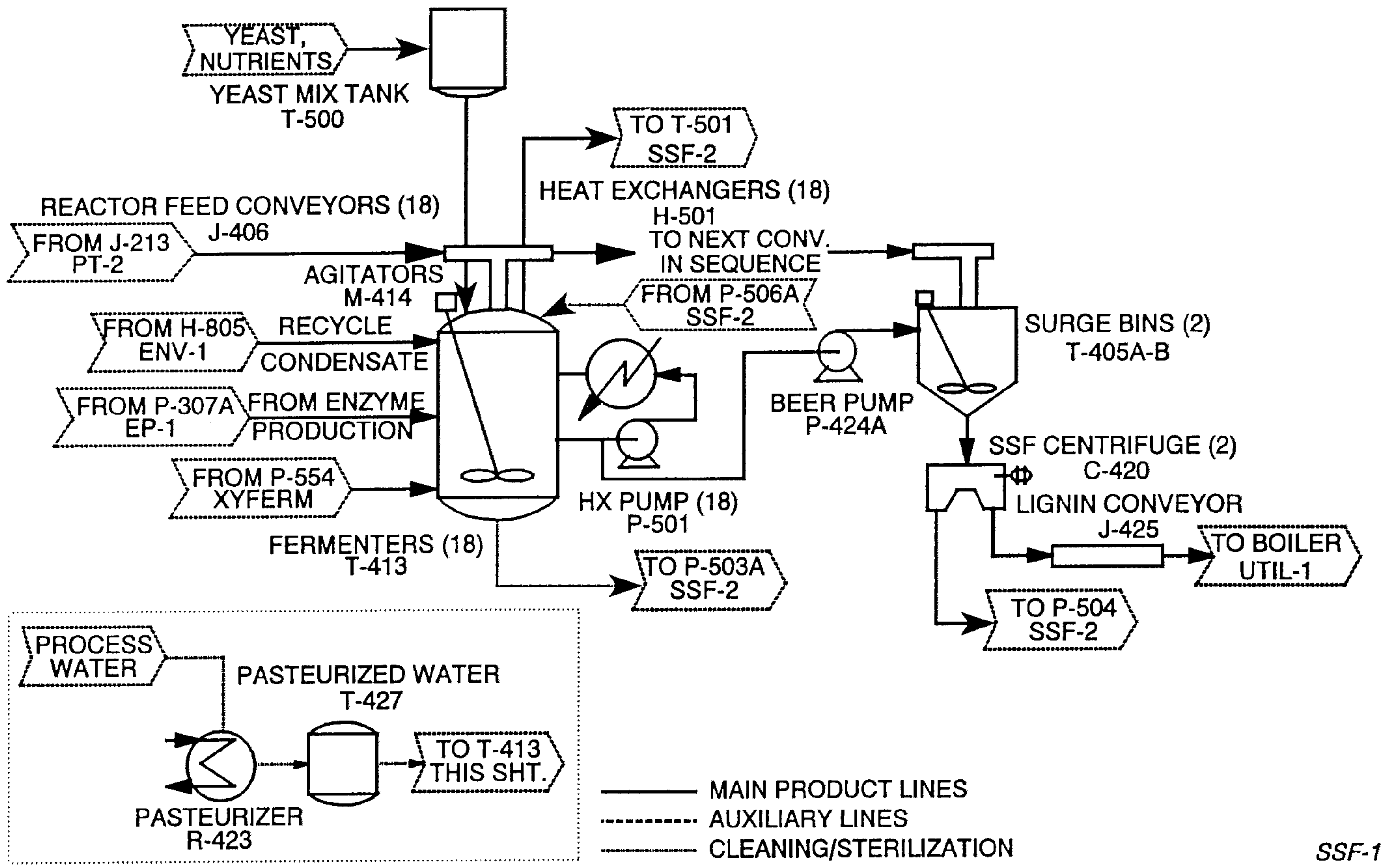
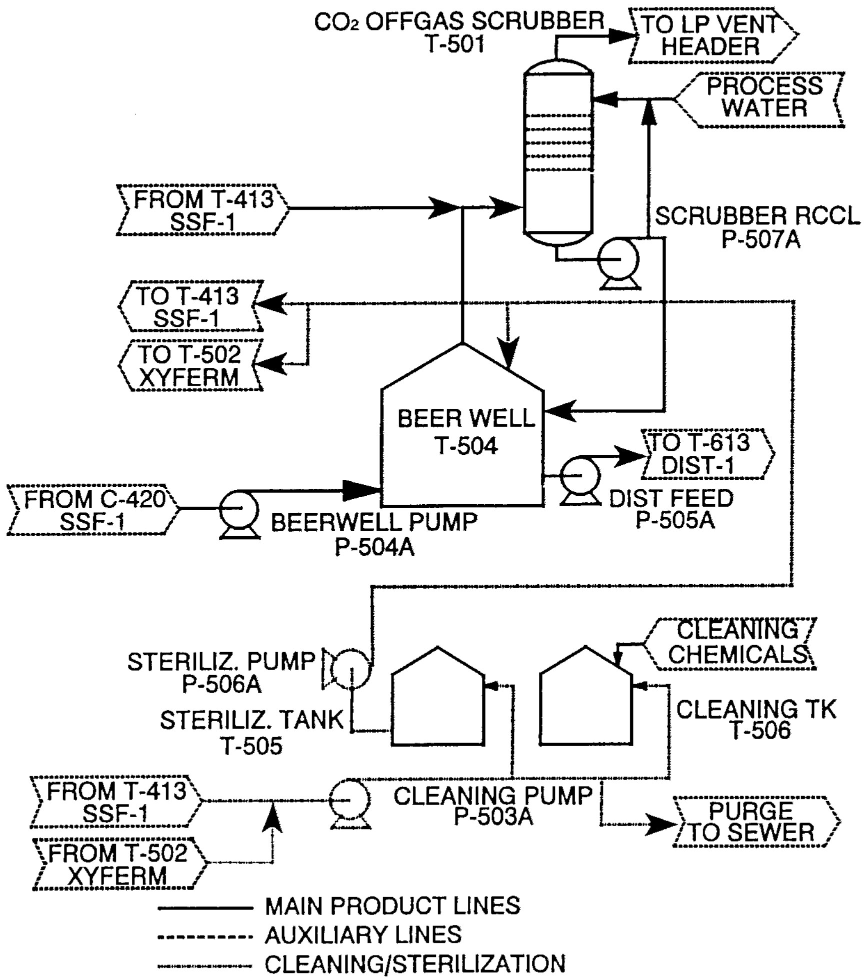


Figure 3.6 Simultaneous saccharification and fermentation, part 1



SSF-2

Figure 3.7 Simultaneous saccharification and fermentation, part 2

The yeast, *Saccharomyces cervisiae*, is manually added to a yeast mix-tank (T-500) and then transferred to the fermenter as it is being filled (about 300 lb of yeast per batch). The yeast is purchased rather than manufactured on location. The inlet mash temperature is about 80°F, gradually rising to a maximum of about 95°F during the peak fermentation period. Cooling is provided during this peak period by recirculation of the mash through the fermenter cooler. Well water at 60°F is the cooling medium. The recirculation for cooling also serves to agitate the tank. At the end of the fermentation period, the fermenter contents are transferred to beer well T-504 (Figure 3.7), from which they are pumped continuously to the distillation section.

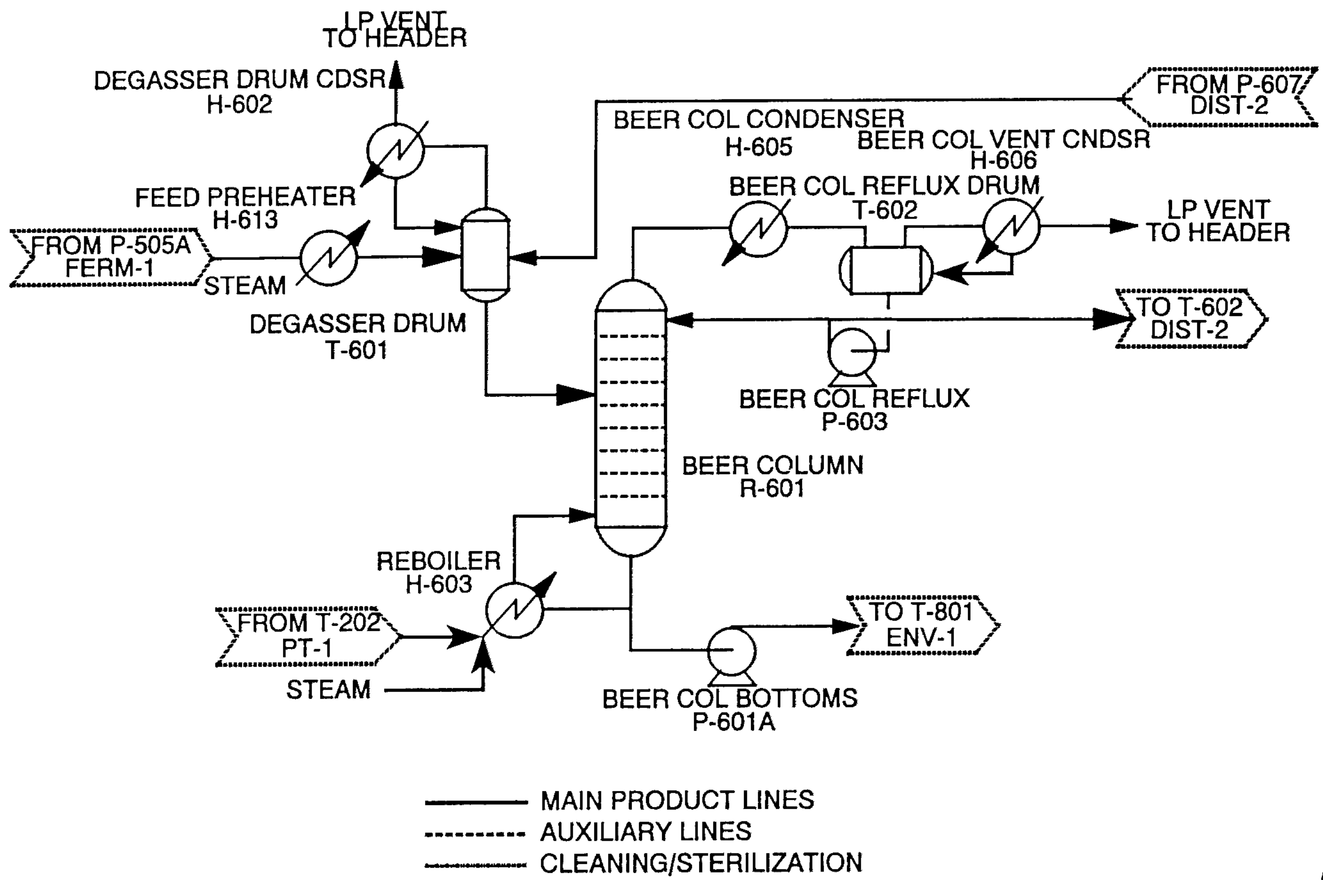
The fermenters are cleaned and sterilized by means of automatic spraying machines which are installed in each fermenter tank. Each tank has two such spraying machines. After each fermentation cycle, the tank is washed with a cleaning solution, sterilized with an iodine solution, and rinsed with clean sterile water in preparation for the next cycle.

3.6 Ethanol Purification (Section 600)

Dilute beer containing 4.8 weight percent ethanol from fermentation (Section 500) is stored in the beer well at 90°F to prevent contamination. From there it is pumped by distillation feed pump P-505A on flow control to feed preheater H-613 (Figure 3.8). The fresh feed along with recycled bottoms from the rectification column (Figure 3.9), then enters degasser drum T-601 where some of the dissolved CO_2 leaves as flash vapor. Degasser drum condenser H-602 uses cooling water to condense and recover the light organics and some water from the flash vapor and these liquids return to T-601. The remaining cooled vent vapor passes into the low pressure vent header.

The degassed feed then enters beer column R-601 on tray twelve. R-601 is an atmospheric column with 16 sieve trays designed to concentrate the ethanol to 40 weight percent in the overhead while losing negligible amounts in the stillage. Heat is supplied by beer column reboilers H-603A,B. These are thermosiphon reboilers whose heating medium is 50 psig steam controlled by the column bottoms pressure, with flash vapors from pretreatment providing additional energy. The flow of stillage from the bottom of the beer column is controlled by the column liquid level. The stillage is pumped by beer column bottoms pump P-601A to first effect evaporator T-801.

Overhead vapor from R-601 is condensed in beer column condenser H-605 using cooling water and collected in beer column reflux drum T-602. Vapor from T-602 is further condensed in beer column vent condenser H-606 using cooling water and the liquid is returned to the drum. The remaining cooled vent vapor passes into the low pressure vent header. Reflux to R-601 is pumped by beer column reflux pump P-603A on flow control reset by reflux drum level control. Distillate product is fed to tray 14 of rectification column



DIST-1

Figure 3.8 Ethanol purification, part 1

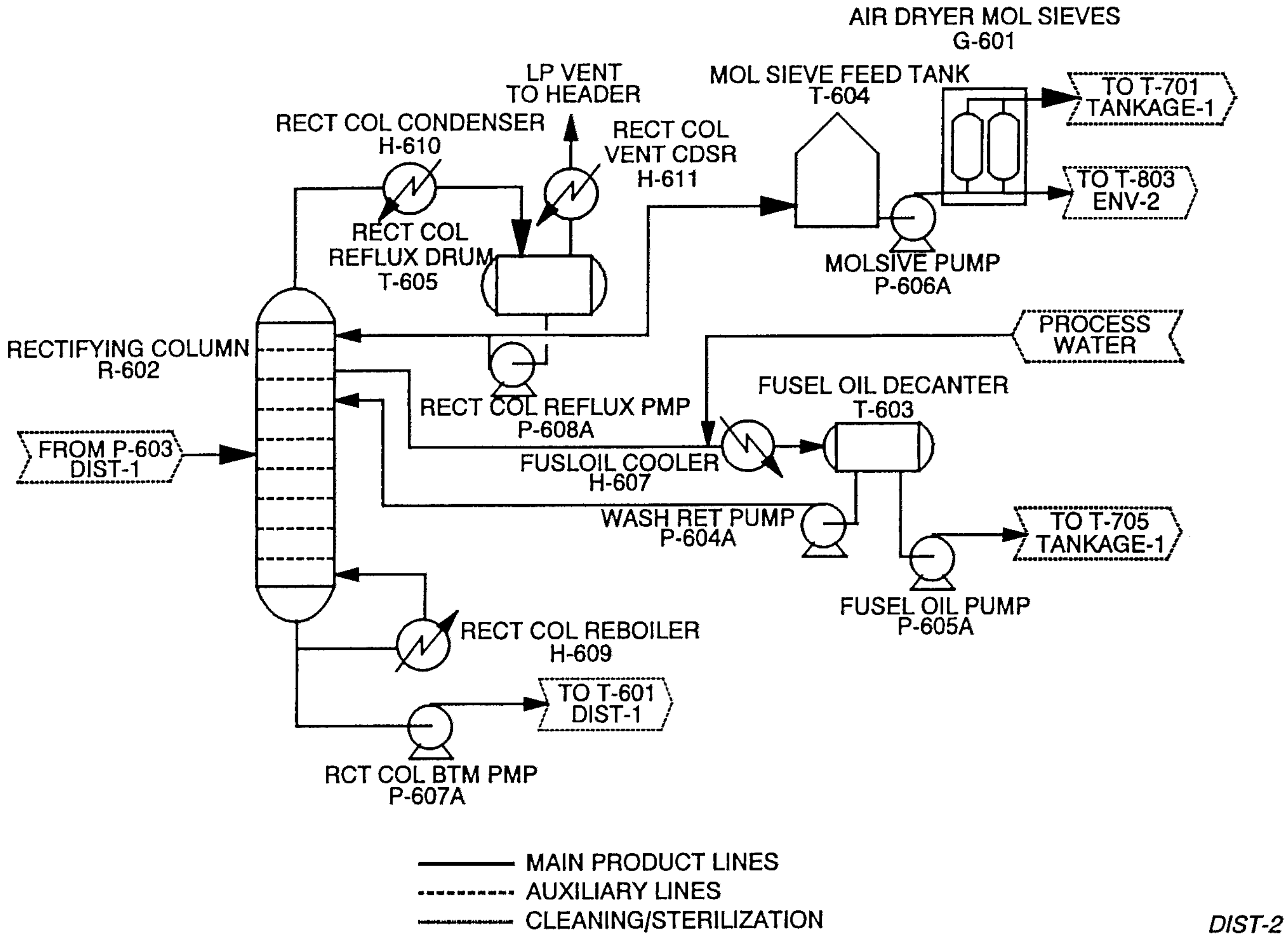


Figure 3.9 Ethanol purification, part 2

Ethanol product from molecular sieve package unit G-601 is pumped to ethanol product tanks T-701A/B. Prior to entering the tank, the ethanol is denatured by blending with gasoline in the proportion of five gallons of gasoline per one hundred gallons of ethanol product. Gasoline delivered by truck or railcar is stored in gasoline storage tank T-710 which has two days capacity. The gasoline is blended into the ethanol by gasoline blending pump P-71OA on flow control. Denatured ethanol is transported to the shipping point by ethanol export pumps P-701A/B.

Fusel oil product is pumped by fusel oil pump P-605A to fusel oil storage tank T-705. Fusel oil is transported to the shipping point by fusel oil export pump P-705A.

Concentrated sulfuric acid delivered by truck or railcar and stored in sulfuric acid storage tank T-703. Moisture from the atmosphere is kept from the acid by desiccant air filter F-703 mounted on the vent of T-703. Sulfuric acid transfer pump P-703A transports acid to sulfuric acid day tank T-202.

Diesel fuel delivered by truck is stored in diesel fuel tank T-709 which has a one month capacity. Fuel is delivered to mobile equipment by diesel fuel pump P-709A. Fire water tank T-707 and fire water pump P-707A are remotely located from the process units and offsite tankage.

3.8 Waste Treatment (Section 800)

The environmental section consists of three parts; an evaporation section, an anaerobic digestion and aerobic treatment section, and a low pressure vent system.

In the evaporation section, a three-stage evaporator is used to produce mixed solvents at 60% concentration in water to be burned as liquid fuels to produce steam (Figure 3.11). About half of the recovered condensate is recycled to theSSF reactors, and about half is sent to anaerobic digestion. Beer column bottoms pump P-601A transports stillage to first effect evaporator T-801 running at 14 psig. Heat for the first effect is provided by first effect heater H-801 using 50 psig steam. Vapors from the first effect are condensed in second effect heater H-802 which provides heat for second effect evaporator T-802 running at 11 psia. Liquid is pumped from the first effect to the second effect on level control by first effect pump P-801A. Similarly, second effect pump P-802A feeds third effect evaporator T-803 on level control and vapors from the second effect condense and provide heat for the third effect in third effect heater H-803. Vapors from the third effect are condensed in evaporator condenser H-804 using cooling water. The pressure in the third effect is controlled at 3 psia by evaporator vacuum system G-804 using 50 psig steam as the motive source and a pressure control bypass which allows steam back to the inlet to H-804. Liquid from the third effect, containing a mixture of about sixty percent organics and forty percent water, is pumped on level control to rotary filter F-804 where precipitated salts are separated and sent to disposal. Solid-free liquid is then

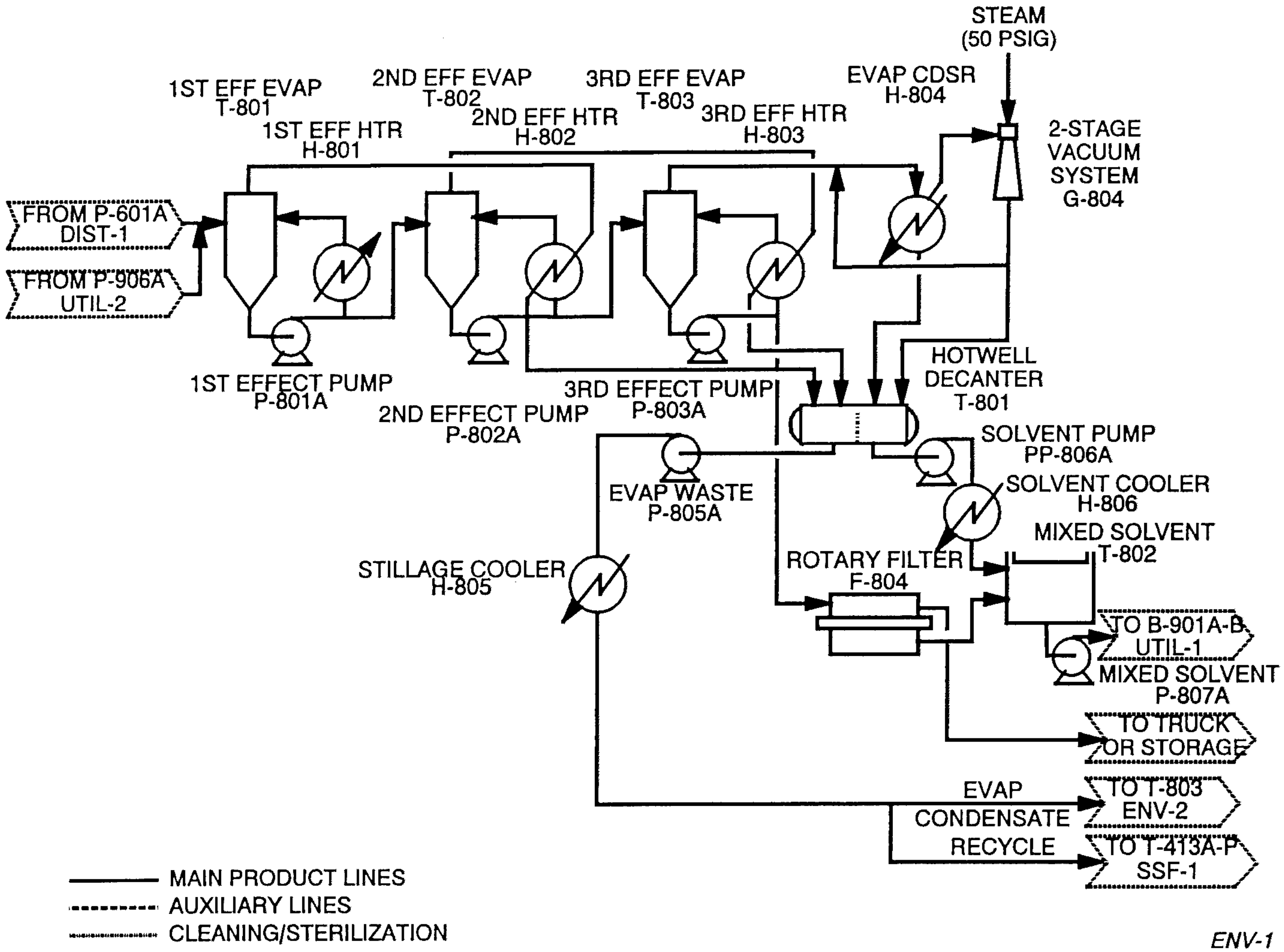


Figure 3.11 Environmental, part 1 (evaporation)

sent to mixed solvent tank T-802 for storage. Boiler feed water demineralizer regeneration effluent from demineralizers G-903A/B and boiler water blowdown from flash drum T-902 are also treated by evaporation.

Condensate from the evaporators is collected in hotwell decanter T-801 where liquid phase separation can occur. The organic layer overflows a baffle and is pumped on level control by solvent pump P-806A to solvent cooler H-806, cooled by cooling water, and sent to T-802. Waste water is pumped on interface level control by evaporator waste water pump P-805A to stillage cooler T-805, cooled and sent to anaerobic treatment. T-802 contains mixed solvents collected from T-801 and F-804. Solvent is transported by mixed solvent pump P-807A to boilers B-901A/B where the solvent is burned to produce high pressure steam.

In the anaerobic treatment section, the Celrobic process is used. Waste water (evaporator condensate from H-805, flume washings from P-101, and molecular sieve runoff from G-601) is collected in equalization-tank T-803 (Figure 3.12). Reactor feed pump P-808A pumps waste water on flow control to anaerobic reactor MT-804, after mixing with recycled reactor effluent. Ninety percent of the process effluent COD is removed in this high rate anaerobic digester. The digester produces methane gas which passes through off-gas suction knock-out pot T-806. A split range pressure controller maintains the reactor outlet pressure by recycling off-gas. Excess system pressure will open a back pressure control valve allowing off-gas to pass to boiler B-901A/B where it is combusted. Liquid effluent from T-804 overflows into reactor surge drum T-805 and is pumped on level control by reactor recycle pump P-809 to biotreater T-807. A portion is recycled to the reactor inlet.

The remaining COD is removed in the aerobic treatment section. Effluent from anaerobic treatment is combined with recycled underflow from the secondary clarifier and filtrate from the belt filter and fed to biotreater T-807. Effluent from the biotreater is combined with sand filter backwash and enters secondary clarifier T-808. Overflow from the clarifier is used as cooling tower makeup and is pumped on level control by final effluent pump P-816A.

A portion of the clarifier underflow is recycled on flow control as feed to the biotreater and the remainder is pumped on flow control by sludge pump P-813A to filter feed tank T-808. Filter feed pump P-814A feeds sludge to belt filter press F-810 which separates solids for disposal. Filtrate is collected and recycled to the biotreater. Low pressure vents from all sections are drawn into low pressure vent knock-out drum T-810 to remove entrained liquids and transported by low pressure vent blower P-817A/S to the firebox of boilers B-901A/B where it is mixed with combustion air. The suction pressure of P-817A is controlled using a pressure control bypass line.

Most of the low pressure process vents contain air or carbon dioxide saturated with water and small amounts of volatile organics. The air comes both from displacement during filling of vessels and from release of air

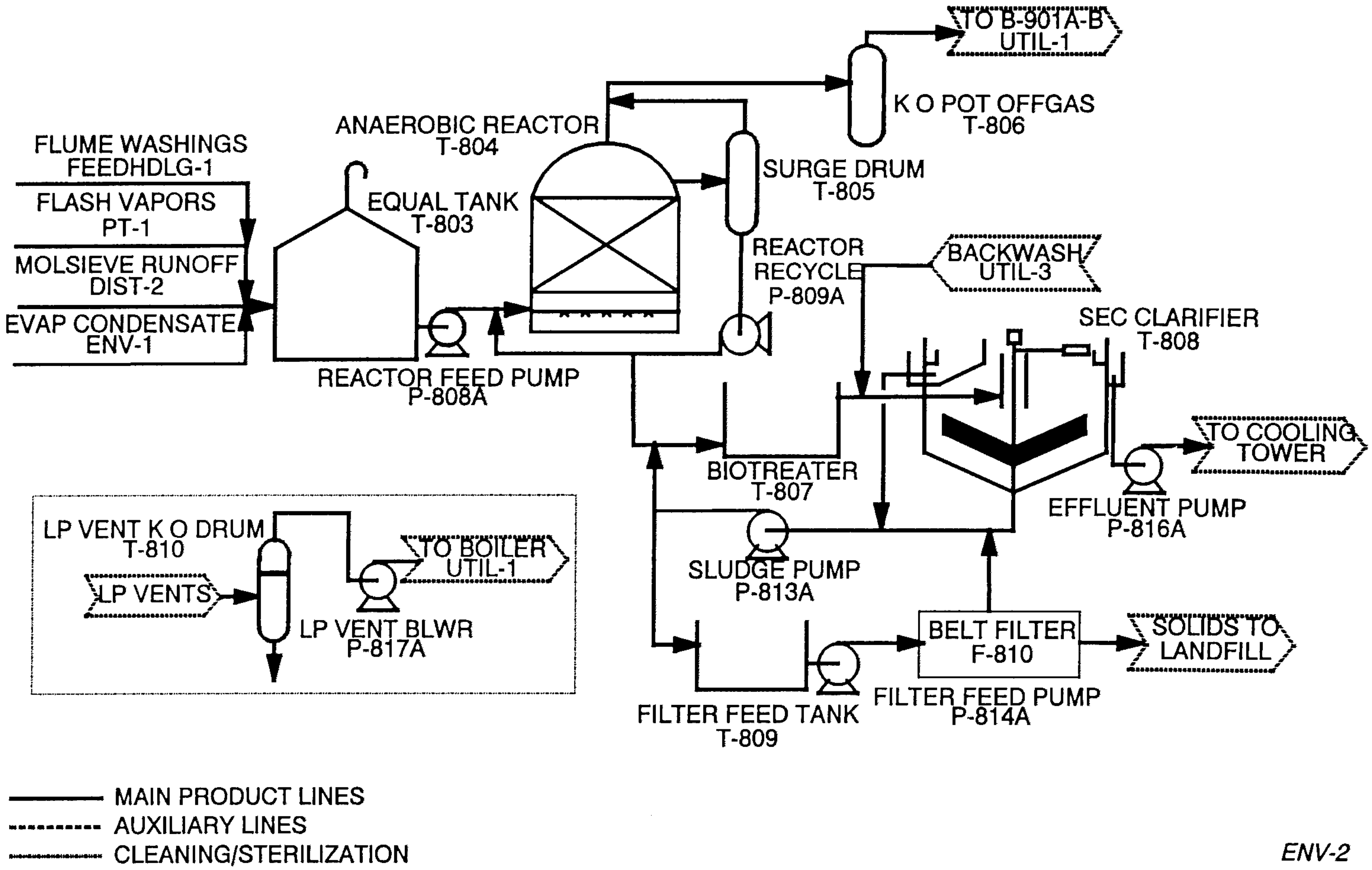


Figure 3.12 Environmental, part 2 (aerobic/anaerobic treatment)

trapped in pores in the wood. The organics are mostly ethanol and the extraneous material from wood such as terpenes (*e.g.,* alpha pinene) and extractive resins (*e.g.,* fatty acids). These would present odor problems if vented directly to the atmosphere. Scrubbing with water would be inadequate since alpha pinene and oleic acid are insoluble. Recovering such small amounts of organics by chilling and condensing is uneconomical. The vents are therefore simply incinerated by mixing with combustion air in the boilers.

3.9 Utilities (Section 900)

The utilities section consists of the boilers and steam distribution system, the boiler feedwater system, the turbogenerator system, the cooling water system, the process water system, and the plant and instrument air system.

Turbogenerator package unit G-911 (Figure 3.13) is rated to take 1100 psia, 825°F steam and generate 42.4 MW of electricity. (This quantity is insufficient to meet the plant's total demand, so additional electric power will be purchased.) The unit includes a condenser, vacuum ejector set, and controls. 7% of the steam is let down in a topping cycle to 150 psig, 48% is let down to 50 psig, and the remainder is condensed at 89 mm Hg. Turbine condensate pump P-910A returns condensate to the boiler feedwater system.

The boilers B-901A/B are designed to burn solid, liquid, and gaseous fuel byproducts from the process sections in order to generate 1100 psia steam for use in the process and for electric power generation (Figure 3.13). Process requirements amount to 54% of the total available steam. Solid fuels, *i.e.* lignin, comprise about 79% of total byproduct fuel heating value. Liquid fuel, comprising about 21% of total fuel heating value, is a mixture of solvents, sugar, acids, dissolved organic salts, and soluble tars in a 35-40% water mixture. Fuel gas comes from anaerobic digestion and comprises less than 1% of total heating value. Gas and liquid fuels are burned directly but lignin from pretreatment is fed to a Flakt type drying system which dries and fluidizes the solids into the burners using boiler flue gas. No other solid fuel (such as coal) is required as there is always an excess of byproduct fuel available over normal process section energy requirements. Diesel fuel is used for startup.

Low pressure vents are collected, compressed, and injected into the boilers to mix with combustion air prior to burning. None of these vents contain an explosive mixture either singly or in combinations. They contain mostly air or carbon dioxide saturated with water and very small amounts of organics. Process users of 1100 psia steam are fed directly from the high pressure steam header. Users of 130 and 50 psig steam are supplied with steam letdown from 1100 psia through a turbogenerator. All recoverable condensate is returned to the boiler feedwater system.

Boiler feedwater preparation consists of makeup process water demineralization, condensate polishing, deaeration, and chemical dosing

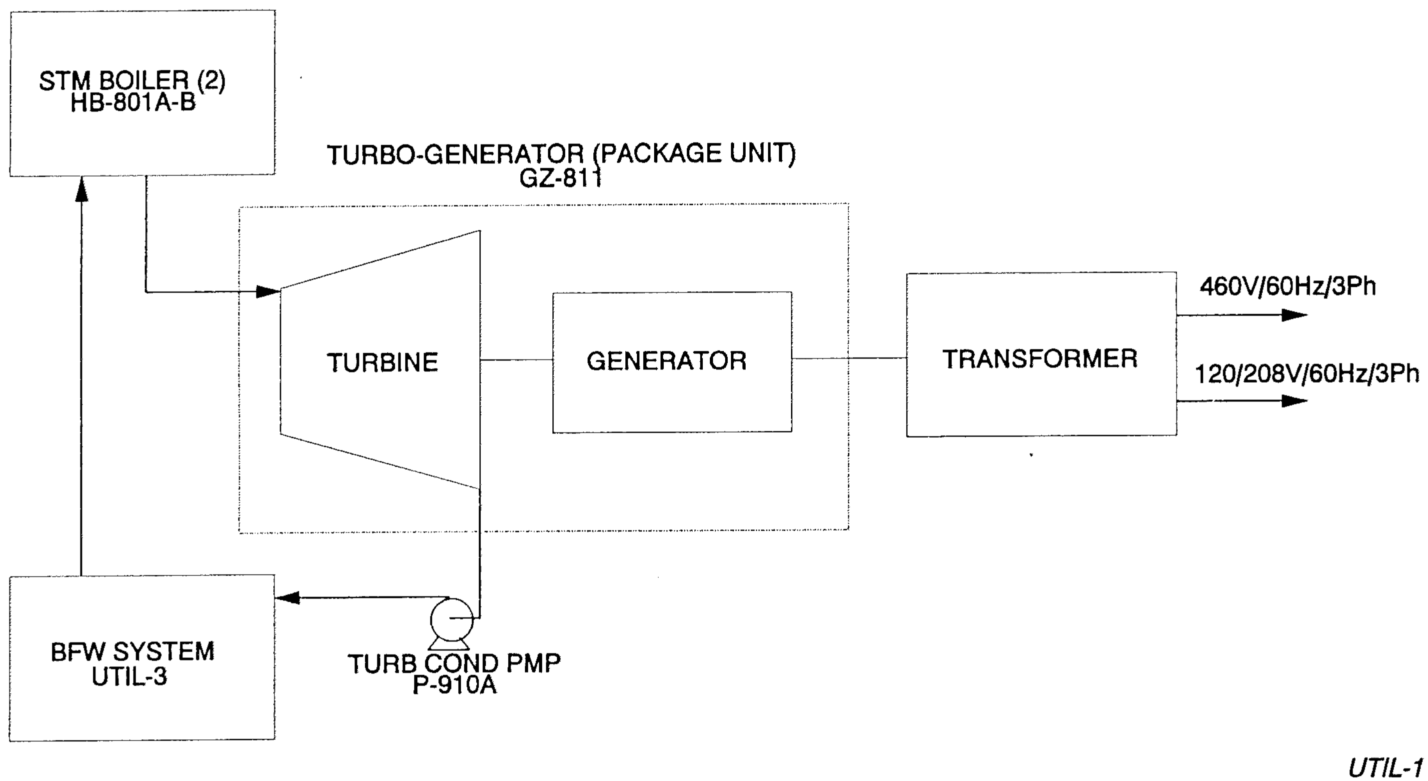


Figure 3.13 Utilities, part 1

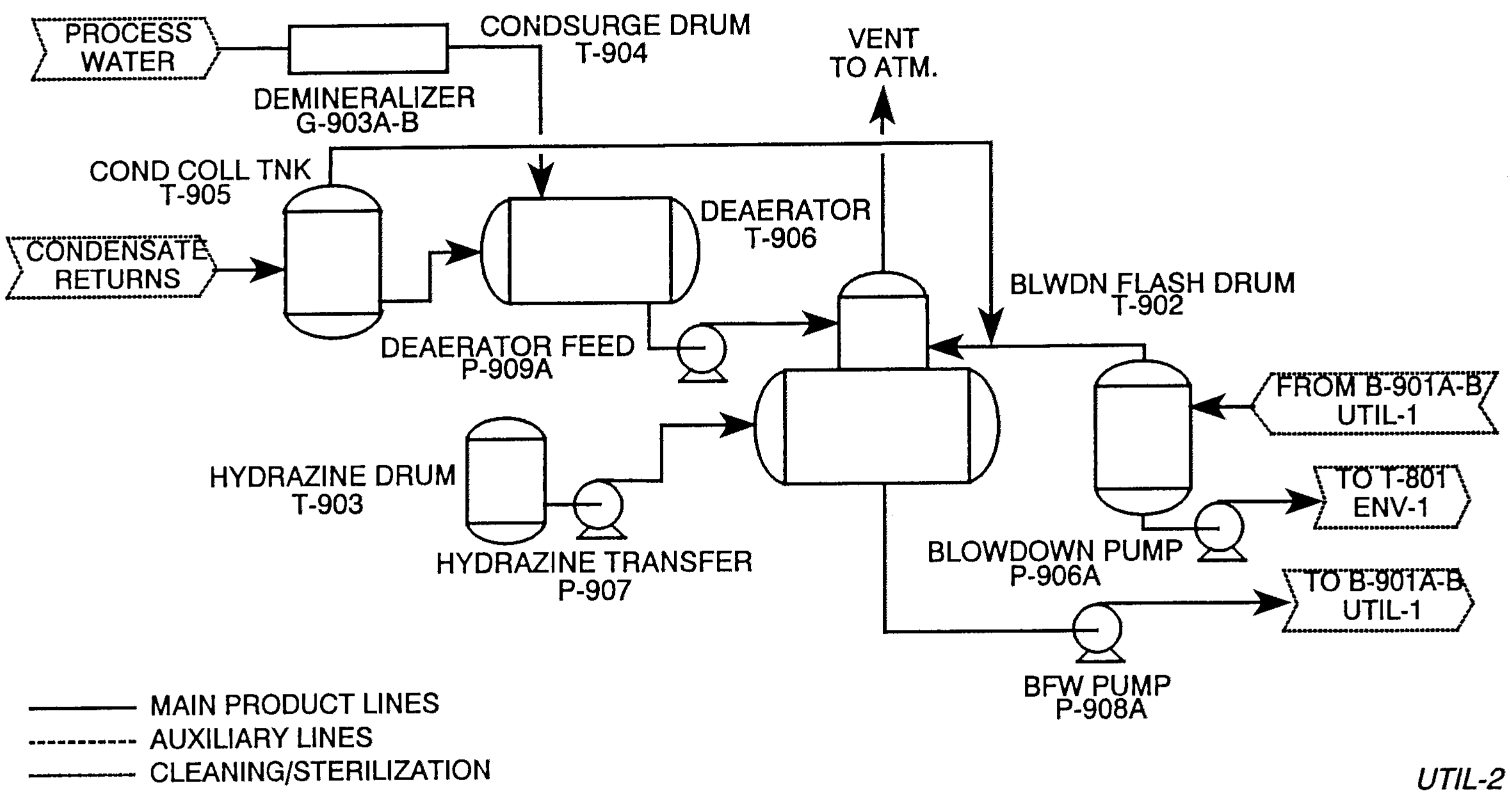


Figure 3.14 Utilities, part 2 (boiler feedwater)

(Figure 3.14). Recoverable condensate is collected in condensate collection tank T-905. Flashed low pressure steam is used in the deaerator. Condensate is then pumped through condensate polisher G-904 to condensate surge drum T-904. Makeup process water is introduced on level control via demineralizers G-903A/B, a package unit consisting of anion and cation resin beds with intermediate degasifier and regeneration equipment. Deaerator T-906 is fed by deaerator feed pump P-909A on flow control reset by level control. The deaerator runs at 10 psig and expels air and steam to the atmosphere. Steam to the deaerator is fed on flow control and is supplied by flashes from recovered condensate and from boiler blowdown flash drum T-901 and made up from the 50-psig steam header. Boiler blowdown is sent to evaporator T-801 by blowdown pump P-906A.

Demineralizer regenerant effluent is sent to evaporator T-801. Deaerated boiler feedwater is dosed with hydrazine made up in hydrazine addition unit G-907. BFW pump P-908A feeds boilers B-901A/B. Hydrazine is transferred from hydrazine drum T-903 by hydrazine transfer pump P-907 to be mixed with condensate in G-907. A shed encloses the hydrazine addition unit.

Process water preparation consists of filtering well water pumped by well water pumps P-913A-B through a sand filter and anthracite bed F-901 (Figure 3.15). Process water transfer pump P-902A feeds process water tank T-901. Backwash feed pumps P-904A/B provide a daily backwash flow to F-901. Backwash overflow is collected in backwash transfer tank T-905 and is transported by backwash transfer pump P-905A to clarifier T-908. Process water is distributed via a process water ring main pressurized to 60 psig by process water circulating pump P-903A.

Cooling tower G-912 provides cooling water at 88°F with an allowable temperature rise of 30°F. Cooling water distribution is maintained at a pressure of 60 psig by cooling water pumps P-912A-H. Cooling tower blowdown containing dissolved salts at 2000 ppm is discharged directly to the environment. Cooling tower blowdown contains zero COD.

Plant and instrument air are provided by air compressor P-911 operating on pressure control from plant air receiver T-906. Instrument air receiver T-907 is fed via instrument air dryer G-910.

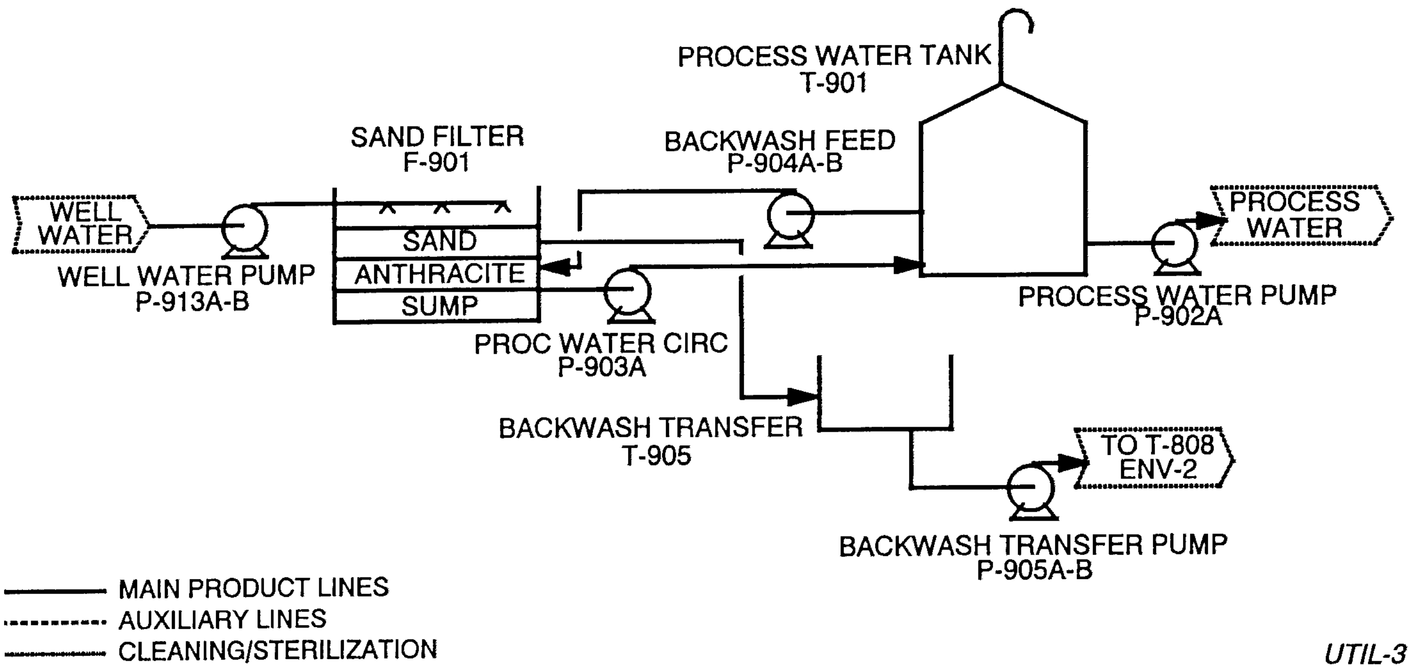


Figure 3.15 Utilities, part 3 (process water)

4 Process Economics

This section presents the economics of the SHF process on two different bases. The first is the basis used in the original FATE study. This case uses a 1985 capital cost basis, and a 12.9% fixed charge rate (FCR). The second case uses capital costs which are inflated to 1990 dollars, and a FCR of 20%. The effect of inflation from 1985 to 1990 on the overall economics is quite small, however, the higher FCR used in more recent studies dramatically increases the estimated cost of ethanol production.

4.1 Original (1985) Basis

4.1.1 Economic Assumptions

The cost estimates presented in this report were originally prepared using 1985 dollars, are based on a facility producing 52 million gallons per year of ethanol (actually 54.4 million gallons of fuel product since gasoline is added as a denaturant), operating for 8000 hours per year. Fixed capital assumptions have been simplified and aggregated into the fixed charge rate (FCR), according to the methods of Argonne National Laboratory's (ANL) Industrial Applications Group (ANL 1985). The FCR accounts for return of the initial capital investment (including working capital), as well as depreciation, taxes, and profits. The value of FCR used in this study is 12.9%, which is roughly equivalent to a 14% discounted cash flow annual rate of return (DCFIRR) over a 10-year economic lifetime for a nonregulated industry with tax preferences.

4.1.2 Facility Capital Costs

Most of the equipment costs were estimated with the COST® system, developed by the ICARUS Corp. This system uses computer models to develop a comprehensive purchased and installed cost estimate for each process section, as opposed to a factored cost estimate where the purchased equipment costs are simply multiplied by installation factors to yield the installed costs. The accuracy of the computer-aided estimate is generally much higher than the factored estimate. An ICARUS cost estimate was prepared for each process section.

The total purchased equipment cost for the facility is $55.0 million (Table 4.1). Total installed cost is $105.7 million. Total capital investment including engineering, contingencies, *etc.*, is $166.0 million. Startup and working capital add another $15.7 million, bringing the total facility cost to $181.6 million, or 3.3 times the purchased equipment cost.

Figure 4.1 shows the capital cost breakdown by process section (excluding startup and working capital). SSF (32.8%) and enzyme production (15.6%) are the most important sections in terms of cost, and therefore should be the focus of potential technological improvements aimed at reducing costs.

Table 4.1 SHF facility master cost summary in $k (1985)

(ALL IN $K)	MATERIAL	MANPOWER	SUBCON.	TOTAL	% OF PURCH.	% OF TOTAL
1. PURCHASED EQUIPMENT	32,447.9	0.0	22,574.2	55,022.1	100.0	33.2
2. EQUIPMENT SETTING	0.0	883.0	0.0	883.0	1.6	0.5
3. PIPING	14,832.3	5,158.0	0.0	19,990.3	36.3	12.0
4. CIVIL	3,161.8	2,498.4	0.0	5,660.2	10.3	3.4
5. STEEL	2,885.4	991.8	0.0	3,877.1	7.0	2.3
6. INSTRUMENTATION	5,522.8	976.9	0.0	6,499.6	11.8	3.9
7. ELECTRICAL	6,291.6	1,726.1	0.0	8,017.7	14.6	4.8
8. INSULATION	2,955.2	1,885.3	0.0	4,840.5	8.8	2.9
9. PAINT	251.1	649.4	0.0	900.5	1.6	0.5
10. UNIT TOTALS	68,348.0	14,768.8	22,574.2	105,691.0	192.1	63.7
11. ENGINEERING			12.96%	13,692.7	24.9	8.3
12. CONSTRUCTION OVERHEAD			15.30%	16,171.6	29.4	9.7
13. BARE PLANT COST				135,555.2	246.4	81.7
14. CONTRACTOR'S FEE			3.00%	4,066.6	7.4	2.5
15. CONTINGENCIES			15.45%	20,943.4	38.1	12.6
16. SPECIAL CHARGES			3.98%	5,396.7	9.8	3.3
17. TOTAL CAPITAL INVESTMENT				165.961.9	301.6	100.0
18. STARTUP (5% OF TCI)			5.00%	8,298.1	15.1	5.0
19. WORKING CAPITAL						
ACCOUNTS RECEIVABLE (1 MO. NET PROD. COST)				3,729.1	6.8	2.02
CASH (1 WK. GROSS PROD. COST)				893.8	1.6	0.5
WAREHOUSE, SPARES (3% OF TCI)			3.00%	4,978.9	9.0	3.0
(ACCOUNTS PAYABLE: 1 MO. RAW MATERIALS)				(2,240.0)	(4.1)	(1.3)
20. OTHER CAPITALIZED COSTS				15,659.9	28.5	9.4
21. TOTAL PROJECT COST				181,621.8	330.1	109.4

Pretreatment (7.3%) is a third section where cost savings might be realized by technological improvement. Although the costs of items in the utilities block (especially the boiler and turbine-generator) are also high, these costs are less sensitive to technological improvements than the aforementioned blocks.

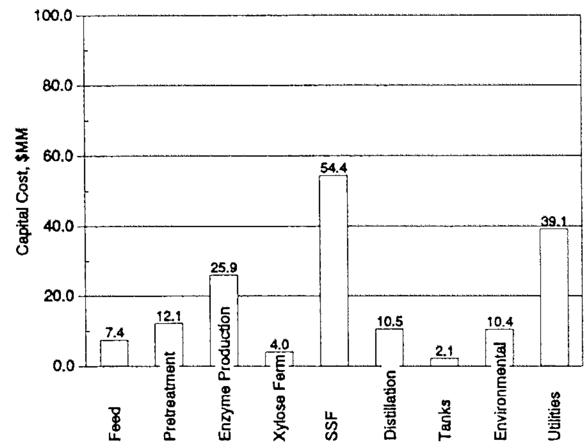


Figure 4.1 Capital cost by section (1985 $)

4.1.3 Annual Operating Costs

Annual operating costs were calculated from the material and energy balance. Operating costs consist of raw materials, utilities, labor, maintenance costs, taxes and insurance, less by-product credits (if any). Maintenance (O&M) costs are taken as $0.040/yr/$ of capital investment, and general overhead is taken as 60% of the total of labor and O&M. Labor costs are estimated as 0.001 man-hr/yr/$ of capital investment (at $15.40/man-hr).

4.1.4 Cost of Production

Table 4.2 shows cost of production summary. The annual revenue requirement is $69.9 MM, yielding a total ethanol product cost of $1.35/gal. The prominent items are feedstock cost ($0.52/gal) and capital cost ($0.45/gal). Both of these items would be expected to decrease substantially with improvements in the hydrolysis and fermentation yields. Labor and

Table 4.2 SHF facility annual operating cost summary (1985 $)

Raw Materials	Units/Gal	Cent/Unit	Annual $MM	Cent/Gal
Wood lb.	24.7	2.1	26.88	51.86
Sulfuric Acid lb.	0.5	3.25	0.85	1.64
Lime lb.	0.5	1.75	0.46	0.88
Chemicals	0.0		0.58	1.12
Utilities				
Water 1000 gal	0.010	75.0	0.38	0.73
Steam	0.000	550.0	0.00	0.00
Elect. (MWhr)	0.001	2920.0	1.73	3.34
Labor hr.	0.003	1540.0	2.71	5.23
Overhead & Maint			12.89	24.88
Ann. Op Cost			46.48	89.67
Capital Charges			23.43	45.20
			69.91	134.87

maintenance costs, being functions of capital investment, would similarly decrease. In other versions of this process, the credit for electricity generated on-site from the combustion of lignin and other organics reduces the product cost, sometimes substantially. But the magnitude of the electricity credit is inversely proportional to the hydrolysis/fermentation yield: the more feedstock which can not be fermented to ethanol, the more will be burned to produce steam and electricity. In this case, where substantial amounts of the xylose are converted to ethanol, the overall product conversion is so high that additional electricity must be purchased to meet plant requirements.

4.2 Inflation-Adjusted Case

Table 4.3 shows the master cost summary for the inflation-adjusted case. Capital equipment has been inflated to 1990 dollars using the *Chemical Engineering* 1990 annual average. The 1985 CE Cost Index is 325, and the 1990 CE Cost Index is 357, an increase of roughly 10%. The FCR has been increased to 20%. Purchased capital equipment cost has increased from $55.0 million to $60.4 million, and the total project cost to $198.3 million (including startup and working capital).

Table 4.4 shows the annual operating cost summary. The ethanol product cost has increased from $1.35/gallon to $1.68/gal, primarily because of the higher cost of capital (FCR).

Table 4.3 SHF facility master cost summary in $k (adjusted to 1990 $)

(ALL IN $K)	MATERIAL	MANPOWER	SUBCON.	TOTAL	% OF PURCH.	% OF TOTAL
1. PURCHASED EQUIPMENT	35,886.9	0.0	24,517.9	60,404.8	100.0	33.3
2. EQUIPMENT SETTING	0.0	960.5	0.0	960.5	1.6	0.5
3. PIPING	16,111.2	5,610.5	0.0	21,721.7	36.0	12.0
4. CIVIL	3,402.5	2,709.2	0.0	6,111.7	10.1	3.4
5. STEEL	3,146.5	1,079.3	0.0	4,225.8	7.0	2.3
6. INSTRUMENTATION	6,015.6	1,062.3	0.0	7,077.9	11.7	3.9
7. ELECTRICAL	6,859.0	1,881.9	0.0	8,740.9	14.5	4.8
8. INSULATION	3,206.6	2,046.1	0.0	5,252.7	8.7	2.9
9. PAINT	273.3	707.4	0.0	980.7	1.6	0.5
10. UNIT TOTALS	74,901.5	16,057.3	24,517.9	115,476.7	191.2	63.7
11. ENGINEERING			12.99%	15,003.4	24.8	8.3
12. CONSTRUCTION OVERHEAD			15.30%	17,669.3	29.3	9.7
13. BARE PLANT COST				148,149.5	245.3	81.7
14. CONTRACTOR'S FEE			3.00%	4,444.5	7.4	2.5
15. CONTINGENCIES			15.45%	22,889.2	37.9	12.6
16. SPECIAL CHARGES			3.98%	5,901.5	9.8	3.3
17. TOTAL CAPITAL INVESTMENT				181,384.6	300.9	100.0
18. STARTUP (5% OF TCI)			5.00%	9,069.2	15.0	5.0
19. WORKING CAPITAL						
ACCOUNTS RECEIVABLE (1 MO. NET PROD. COST)				3,729.1	6.2	2.1
CASH (1 WK. GROSS PROD. COST)				906.1	1.5	0.5
WAREHOUSE, SPARES (3% OF TCI)			3.00%	5,441.3	9.0	3.0
(ACCOUNTS PAYABLE: 1 MO. RAW MATERIALS)				(2,240.0)	(3.7)	(1.2)
20. OTHER CAPITALIZED COSTS				16,906.0	28.0	9.3
21. TOTAL PROJECT COST				198,290.6	328.3	109.3

Table 4.4 SHF facility annual operating cost summary (adjusted to 1990 $)

Raw Materials	Units/Gal	Cent/Unit	Annual $MM	Cent/Gal
Wood lb.	24.7	2.1	26.88	51.86
Sulfuric Acid lb.	0.5	3.25	0.85	1.64
Lime lb.	0.5	1.75	0.46	0.88
Chemicals	0.0		0.58	1.12
Utilities				
Water 1000 gal	0.010	75.0	0.38	0.73
Steam	0.000	550.0	0.00	0.00
Elect. (MWhr)	0.001	4000.0	2.37	4.57
Labor hr.	0.003	1540.0	2.71	5.23
Overhead & Maint			12.89	24.88
Ann. Op Cost			47.12	90.90
Capital Charges			39.85	76.88
			86.97	167.78

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Appendix A: Spreadsheet Printout

A	B	C	D	E	F	G	H	I	J	K	L
	SIMULTANEOUS SACCHARIFICATION AND FERMENTATION				08-Sep-93						
2	/w xylose fermentation and SSF in series								^		
3			————	————	CO2=	28931			| EOH=	42634	
4	|	——|	Enzyme Production			^			|		
5	|		————	————		|			|		
6	|		|			|			|		
7	|		————	————		————	————		———		
8	|	|	Hydrolysis			Fermentation	|		¦	Stillage	
9	————	——>|	Eff CC	90.0% |——>		C6 Eff=	90% |——		Distillation	|——	——>
10	Wood=	160000	————	————		C5 Eff=	70% |		|	Water=	796046
11	Water=	320000	|	|Water=	796046	————	————		———	Solubles=	21100
12	%solid>SSF	12.12%	|	|Gluc=	59043					Total=	817800
13	%C6 > SSF=	6.86%	|	|Xyl =	11243		Water=	796046			
14			|	|Total=	937914		Ethanol=	42634			
15			^	| % C6	7.4%		Total=	860434			
16				|% Solid	4.9%		% EtOH C5	3.77%			
17		Water=	650249				% EtOH SSF	5.01%			Water to:
18		Acid=	3263	Lig/Cel/O=	0						Recycle
19		Steam=	320000	Water=	0						Cooling Twr.
20		Caustic=	3263								
21	Assumptions					Prehydrolysis/Pretreatment					
22			Wet	Dry							
23	Wood Composition		Mixed Hardwood			Hyd Temp (C)		160			
24	Water		0.5			Impreg/flash temp (c)		100			
25	Hexose		0.231	0.462		Water to Imp (C)		85		T (K)	433
26	Xylan		0.12	0.24		Ambient (C)		20			
27	Lignin		0.12	0.24		Acid Concentration (wt%)		1.1%		R=	1.987
28	Other		0.028	0.056		Wt% solids after spray		35.0%			
29	Ash		0.001	0.002		Reaction time		10.0000		Xf (Xylan)	0.83
30			————	————				min			
31			1.000	1.000		Kinetic Parameters					
32							K	1/K	P1	N	E
33	Dry wood feed rate (lb/hr)			160000	MULT	CC hyd	0.00434080541	230.371994403	1.730E+19	1.34	42900
34						Gluc deg	0.00756055363	132.265446316	1.790E+14	0.84	32506
35	f Hexose as nCH2O		14.4%			C5 hyd (f)	0.96517945077	1.03607676189	1.316E+14	1.17	28000
36						C5 hyd (s)	0.08007973709	12.4875534848	3.812E+17	1.17	37000
37						C5 deg	0.05013131236	19.9476126391	2.327E+12	0.69	27130
38											
39											
40								mole f	wt f		
41						CC remaining		0.958	0.95752059012	Hex remain	0.81963762514
42						Glucose formed frm CC		0.041	0.04540133079	weighted Y	0.1685263873
43						Glucose formed from nCH2O		0.927	1.02917188019		
44						HMF formed		0.012	0.00923207029		
45						Xylan remaining		0.076	0.07637841761		
46						Xylose formed		0.602	0.68571808132		
47						Furfural formed		0.322	0.23514358026		
48						Prehydrolysis tank at 161 C by heat exchange w/ products					
49						Kinetics of reaction calculated by parameters of Kwartung					
50						Products flashed in two stages to 100 C					
51						All furfural goes overhead in flash					
52						Equal weights of Calcium Hydroxide and H2SO4 are used in neutralization					
53						1.0%	soluble products lost in lignin wash				
54											
55						Weight fraction solids in centrifuge output				0.35	%SC
56						Lb wash water to cent/lb solids				2.7	W/S
57											
58											
59											
60											
61		Feed 1	>Hyd 2	H2O>Imp	Steam>Imp	Steam>React	H2SO4>Imp	> Reactor	Reactor >	Flash Vap 7	Pretreat >
62	Water	160000	160000	137143	34116	41306		338448	337377	42116	295260
63	Hexose	73920	73920					73920	60588	0	60588
64	Xylan	38400	38400					38400	2933	0	2933
65	Sol solids	8960	8960					8960	8960	0	8960
66	Ash	320	320					320	320	0	320
67	Lignin	38400	38400					38400	38400	0	38400
68	Lime/Gypsum		0					0		0	0
69	Glucose		0					0	13828	0	13828
70	Xylose		0					0	26332	0	26332
71	Furfural		0					0	9030	9030	0
72	Degradation		0					0	682	0	682
73	Ethanol		0					0		0	0
74	H2SO4		0				3263	3263	3263	0	3263
75	Cell Mass		0					0		0	0
76	Enzyme		0					0		0	0
77	Nutrient		0					0		0	0
78	Solvent		0					0		0	0
79	CO2/Air		0					0		0	0
80	Total	320000	320000	137143	34116	41306	3263	501711	501711	51146	450565
81	Capital Costs					Utilities		Actual			
82		Base	Exponent	Actual	Icarus		Electricity	Steam			
83	Section					Section	(MBtu/hr)	(MBtu/hr)			(MBtu/hr)
84	Feed		0.8	10.3	7.4	Feed	1.1	2.0	Steam Produced (85%)		606.2
85	Pretreatment	23.0	0.8	24.9	12.1	Pretreatment	32.8	45.1	HP Steam to proc		45.1
86	Enzyme Prod		0.6	16.9	25.9	Enzyme Prod	3.0	9.9	50 Psig Steam to Proc		294.5
87	Xylose Fermentation		0.6	4.8	4.0	Xylose Ferm	4	0	Condensing Steam > elect		268.6
88	SSF		0.6	51.5	54.4	SSF	96.1	20.8	Purchased Steam		0.0
89	Distillation	13.6	0.6	14.9	10.5	Distillation	12.6	56.5			
90	Offsite Tank	5.1	0.5	5.2	2.1	Offsite Tank	0.8	0.0	Electricity Produced		
91	Environment	11.6	0.6	12.4	10.4	Environment	0.7	185.8		50 psig	58.9
92	Utilities	39.5	0.6	36.3	39.1	Utilities	14.8	9.6		Condensing	86.0
93		————		————	————	Misc.	4.0	11.0			————
94				176.1	166.0		————	————		Total	144.9
96	Including Startup/Working Capital				181.6		170.1	339.6		Export (MW)	-7.4

A	M	N	O	P	Q	R
2						
3						
4						
5						
6						
7						
8	Waste					
9	Treatment		———>	Utilities		
10						
11		Water =	79332			
12		Organics =	79332			
13		Total =	158664			
14						
15						
16						
17			Steam =	608212.6		
18	295289		Elect (MW) =	42.4		
19	295289					
20						
21	Enzyme Production					
22						
23	Productivity (IFPU/l/hr)			83	EPP	
24	Enzyme Titre (FPU/ml)			24.0	ET	
25	Enzyme yield (IFPU/g cell)			250	YENZ	
26	Enzyme Activity (IFPU/gm)			630.0	EA	
27	Neutrients remaining			0%	NR	
28	Temperature C			28	TEP	
29	pH			5	PHEP	
30						
31	lb O2/lb Biomass			0.89	O2/B	
32	Btu/ lb Biomass			5906	BTU/B	
33	lb O2/lb Enzyme			0.5	O2/E	
34	Btu/ lb Enzyme			2950	BTU/E	
35	lb E/ lb Sub			0.5	E/S	
36	Final Cell Concentration (g/L)			13	BCEP	
37	% fungal Growth in Seed Vessels			10.0%	FSV	
38	Residence Time in seed vessels (hr)			156	RTSV	
39	Fermenter Height is 60 ft, active height is 51 ft (aeration calc)					
40						
41	Agitation (hp/1000 gal)					
42	Maximum			7.5	AGEPMAX	
43	Average			3.5	AGEPAVG	
44	VVM air to fermenters			0.2	VVM	
45	Nutrients			g/L	Cent/lb	C/lb EP Broth
46	Corn Steep Liquor			15	11.3	0.170
47	Sodium sulfate (monobasic)			1.4	3.6	0.005
48	Potassium Phosphate (monobasic)			2	2.1	0.004
49	Magnesium Sulfate*7H2O			0.3	0.7	0.000
50	Calcium Chloride*2H2O			0.4	4.7	0.002
51	Tween 80			0.2	73	0.015
52	Antifoam			1	25	0.025
53				————		————
54				20.3		0.221
55				N/L		CN/LBEPB
56	Ammonium Hydroxide			20	10	0.200
57						————
58						0.421
59						CC/LBEPB
60						
61	Water>Cent	Cent>Solids	Cent>Liq	Lime>	Neut >	Wash>Cent
62	276049	189875	381435		381435	8809
63		60588	0		0	
64		2933	0		0	
65			8960		8960	
66		320			0	
67		38400	0		0	
68		0	0	3263	3263	
69			13828		13828	
70			26332		26332	
71			0		0	
72			682		682	
73			0		0	
74			3263		0	
75		0	0		0	
76			0		0	
77			0		0	
78			0		0	
79			0		0	
80	276049	292116	434499	3263	434499	8809
81					Base Case	
82					Electricity	Steam
83			Section		(MBtu/hr)	(MBtu/hr)
84	79.42%	0.6%	Feed		0.5	2.0
85	solids	16.9%	Pretreatment		3.7	87.0
86		2.2%	Enzyme Prod		65.5	5.0
87		1.5%	Hydrolysis		20.0	5
88		36.4%	Fermentation		2.0	5.0
89		11.0%	Distillation		9.5	42.0
90	(MW)	0.3%	Offsite Tank		0.8	0.0
91	17.3	22.1%	Environment		0.7	182.0
92	25.2	6.3%	Utilities		17.0	11.0
93		2.7%	Misc.		4.0	11.0
94	42.4	————			————	————
95	-7.4	100.0%			123.7	360.0

A	S	T	U	V	W	X	Y	Z	AA	AB	AC	AD
	Raw Materials		Units/gal	Cent/Unit	A Cost MM $	Cent/Gal.	Inputs		lb/hr	MM lb/yr	gal/hr	gal/yr
2	Wood lb.		24.7	2.1	26.86	51.86	Wood		160000	1280		
3	Sulfuric Acid lb.		0.6	3.26	0.85	1.64	Sulfuric Acid		3263	26		
4	Lime lb.		0.5	1.75	0.46	0.88	Lime		3263	26		
5	Chemicals		0.0		0.58	1.12	Water		527478	4220		
6	Utilities											
7	Water 1000 gal		0.010	75	0.38	0.73	Outputs					
8	Steam		0.000	550	0.00	0.00	Ethanol		42634		6479	51834599
9							Furfural		0			
10	Labor hr.		0.003	1540	2.71	5.23	Lignin		0			
11	Overhead & Maint				12.89	24.88	Elect. (MW)		-7			
12	Byproducts						CO2		44739			
13	Furfural lb.		0.000	15	0.00	0.00	C6 (Pretreat>)		81080	649		
14	Lignin lb.		0.000	0	0.00	0.00	C5 (Pretreat>)		29675	237		
15	Elect. (MWhr)		-0.001	2920	1.73	3.34	Input Data					
16												
17	Ann. Op Cost				46.48	89.67	Operating rate (hr/yr)		8000			
18	Capital Charges				23.43	45.20						
19												
20					69.91	134.87						
21							Hydrolysis					
22	Fraction of feed to Hyd		0.973	(actually a cellulose balance)								
23	Fraction of feed to EP		0.0272	FFEP			% C6 in reactor (as cell)		10.00%	%SH		
24			FPU/hr	lb E/hr	l broth/hr	~lb broth/hr	Enzyme loading (FPU/gm cellulo		7	ELH		
25	Total Enzyme production		1.87E+06	655	7804	17193	Temperature (C)		37	TH		
26			FPU/HR	LBE/HR		LBEPB/HR	pH		4.8	PHH		
27	Residence time						Residence time (hr)		168.00	RTH		
28	Total Biomass Production		223.8	BEP		broth=E+H2O	Yield (molar)		Hydrolysis			
29	Production in Fed Batch		201.4	BFB			C6 sugars		90%	YC6		
30	Production in Seed Vessels		22.4	BSV			C5 sugars		100%	YC5		
31												
32							Active Volume (ft^3)		2525152	(K gal)	18888	
33	Residence time (hr)		252	RTEP	10.5	Days	Quantity at 454000 gal		29	NFERM		
34	Active Ferm Vol (ft3)		79572	AVFBEP			Hydrolysis Agitation Requirements (hp/1000 gal)					
35	Cellulose Digested		1485				Maximum		10			
36	Cellulose remaining		165				Average		2			
37	Oxygen Required (lb/hr)		507				Residence time (hr)		48	RTX		
38	Heat Evolved (MBtu)		3.1				Active Volume (ft^3)		333836	14	@ 225,000 gal	
39	Compression (hp)		37	CPFBEP			Agitation Power Requirements (hp)				37776	
40	Agitation (hp)		1080	AGFBEP								
41							Fermentation					
42							Glucose					
43	Seed System									% theory	wt. %	
44	Seed Active volume (ft^3)		4303	AVSSEP			CO2 evolution				49.0%	
45	Oxygen Required (lb/hr)		20				Fermentation efficiency			90.0%	45.9%	
46	Heat Evolved (MBtu/hr)		0.1				Conversion to cell mass			10.0%	5.1%	
47	Compression (hp)		1	CPSSEP			Unconverted			-0.0%	-0.0%	
48	Agitation (hp)		42	AGSSEP			Max ethanol conc (%GF)		4.50%			
49										100.0%	100.0%	
50	Energy Consumption						Xylose					
51	Electrical		hp	MBtu			CO2 evolution				34.3%	
52	Agitation		1123	3			Fermentation efficiency (XY)			70.0%	35.7%	
53	Compression		38	0			Conversion to other solubles			0.0%	0.0%	
54	Pumps, etc		0	0			Unconverted			30.0%	30.0%	
55			1161	3								
56	Steam Consumption									100.0%	100.0%	
57	Cooling Water			9			Capital Cost ($/Ann Gal)		0.25			
58							Maximum Ethanol Conc		65.00%			
59							Distillation					
60							Ethanol Recovery Eff		99.5%			
61	Cent>Solids	CentLiq>Xfrm	Dil H2O	Ferm>	CO2>	> Hydrolysis	> Enz Prod	H2O > EP	Nutrient>EP	>Fermenter	Air > Ferm	EP>Hyd
62	6059	384185	0	384185		568888	5172	11366		16538		16538
63		0		0		58937	1650			1650		165
64		0		0		2853	80			80		80
65		8960		8960		8960	0			0		0
66		0		0		311	9			9		9
67		0		0		37354	1046			1046		1046
68	3263	0		0		0	0			0		0
69		13828		-0		-0	0			0		0
70		26332		7899		7899	0			0		0
71		0		0		0	0			0		0
72		682		682		682	0			0		0
73		0		15747		15747	0			0		0
74		0		0		0	0			0		0
75		0		705		705	0			0		201
76		0		0		0	0			0		655
77		0		0		0	0		349	349		0
78		0		0		0	0			0		0
79		0		0	15807	0	0			0	4416	0
80	9322	433968	0	418179	15807	702338	11835	11366	349	19672	4416	17193
81	Cost Summary		Cents/Gallon				Pretreatment Cost Estimates					
82		Capital	Operating	Feedstock	Energy	Total						
83	Section						Equipment		Equipment	Multipliers	Total Cost	Scaled Cost
84	Feed	335.21	255.66		0.09	590.97						
85	Pretreatment	547.69	417.72		2.61	968.01	Feed Handling		0.808	5.184	4.189	10.262
86	Enzyme Prod	1171.47	893.48		0.34	2065.29						
87	Xylose Ferm	179.76	137.10		0.24	317.10	Knife Mills		0.387	2.016	0.780	1.911
88	SSF	2457.86	1874.59		5.62	4338.06	Impregnator & Reactor		2.046	2.016	4.123	8.092
89	Distillation	475.81	362.90		1.70	840.40	Neutralization		2.634	2.016	5.310	3.896
90	Offsite Tank	96.90	73.15		0.04	169.09	Storage		0.111	5.508	0.611	0.840
91	Environment	471.18	359.37		3.42	833.97	Electrical		0.555	1.800	0.999	0.999
92	Utilities	1766.57	1347.36		0.98	3114.90	Separation		4.960	2.016	9.979	9.136
93	Misc.	0.00	0.00		0.42	0.42						
94	Feedstock			36.40	3.34	36.40			10.682		21.803	24.873
95		7601.46	5721.32	36.40	18.79	13277.96						

A	AE	AF	AG	AH	AI	AJ	AK	AL
	Raw Materials		Units/gal	Cent/Unit	A Cost MM $	Cent/lb.		
2	Wood lb.		1.97	2.1	26.88	4.14		
3	Sulfuric Acid lb.		0.04	3.25	0.85	0.13		
4	Lime lb.		0.04	1.75	0.48	0.07		
5								
6	Utilities							
7	Water 1000 gal		0.0002	75	0.10	0.02		
8	Steam		0.0006	550	1.98	0.31		
9	Electricity		0.0001	2920	2.24	0.35		
10								
11	Labor hr.		0.000	1540	0.54	0.08		
12	Overhead & Maint				2.57	0.40		
13	Byproducts							
14	Furfural lb.		0.000	15	0.00	0.00		
15	Lignin lb.		0.000	0	0.00	0.00		
16								
17	Ann. Op Cost				36.62	5.49		
18	Capital Charges				4.53	0.70		
19								
20	BASIS: GLUCOSE				40.16	6.19		
21	Raw Materials		Units/gal	Cent/Unit	A Cost MM $	Cent/lb.		
22	Wood lb.		1.44	2.1	26.88	3.03		
23	Sulfuric Acid lb.		0.03	3.25	0.85	0.10		
24	Lime lb.		0.03	1.75	0.48	0.05		
25								
26	Utilities							
27	Water 1000 gal		0.0001	75	0.10	0.01		
28	Steam		0.0004	550	1.98	0.22		
29	Electricity		0.0001	2920	2.24	0.25		
30								
31	Labor hr.		0.000	1540	0.54	0.06		
32	Overhead & Maint				2.57	0.29		
33	Byproducts							
34	Furfural lb.		0.000	15	0.00	0.00		
35	Lignin lb.		0.000	0	0.00	0.00		
36								
37	Ann. Op Cost				36.62	4.02		
38	Capital Charges				4.53	0.51		
39								
40	BASIS: GLUCOSE & XYLOSE				40.16	4.53		
41	Operating Cost Assumptions							
42								
43	Labor & Overhead		0.001	(manhr/yr) / $ Capital Investment				
44	Maintenance		0.040	* Capital Investment				
45	General Overhead		0.600	* (L&O+M)				
46								
47	Economic Assumptions							
48	ANL Uniform Cost Estimation Assumptions				FCR=	0.129		
49	Environmental							
50		% Water in Evaporated solids			50%	WFWES		
51								
52								
53								
54								
55								
56								
57								
58								
59								

A	Feed > Hyd	Water>Hyd 3	> SSF React	After Hyd	SSFR>	CO2	Wash Water	Solids
60								
62	568888	216882	802308	796046	796046		97096	97096
63	58937		59102	5910	5910			5910
64	2853		2933	0	0			0
65	8960		8960	8960	8960			0
66	311		320	320	320			320
67	37354		38400	38400	38400			38400
68	0		0	0	0			0
69	-0		-0	59043	-0			
70	7899		7899	11243	11243			
71	0		0	0	0			
72	682		682	682	682			
73	15747		15747	15747	42848			
74	0		0	0	0			
75	705		907	907	3918			3918
76	0		655	655	656			
77	0		0	0	0			
78	0		0	0	0			
79	0		0	0	0	28931		
80	702338	216882	937914	937914	908982	28931	97096	145644

A		Elect.	Steam
81	Pretreatment Energy Balance	Elect.	Steam
82		(MBtu/hr)	(MBtu/hr)
83	Feed Handling	1.06	
84			
85	Knife Mills	26.91	
86	Impregnator & Reactor	1.49	
87	Heat H2O		16.06
88	Heat Acid		0.38
89	Heat Feed		28.63
90	Imp-Pre		41.31
91	Available from Flash		41.31
92	Neutralization	0.77	
93	Storage		
94	Electrical		
96	Separation	3.59	
		32.76	45.06

A	AM	AN	AO	AP	AQ	AR	AS	AT	AU
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59									
60									
61	Beer 15	Ethanol 16	Stillage 17	Recycle > Proc	> Environ	Evap >	> Boiler 18	Recycle 19	Cooling 20
62	796048		796048	265349	669909	590577	79332	296289	296289
63	0		0		5910		5910		
64	0		0		0		0		
65	8960		8960		8960		8960		
66	0		0		320		320		
67	0		0		38400		38400		
68	0		0		0		0		
69	-0		-0		-0		-0		
70	11243		11243		11243		11243		
71	0		0		9030		9030		
72	682		682		682		682		
73	42848	42634	214		214		214		
74	0		0		0		0		
75	0		0		3918		3918		
76	655		655		655		655		
77	0		0		0		0		
78	0		0		0		0		
79	0		0		0		0		
80	860434	42634	817800	265349	749242	590577	158664	296289	296289
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Appendix B: Purchased Equipment List

SIMULTANEOUS SACCHARIFICATION AND FERMENTATION, DILUTE ACID PRETREATMENT

CAPITAL EQUIPMENT LIST			BASE		EQUIPMENT COST ($k)		ADJUSTED	DESIGN DATA
P&ID NO.	ITEM	ICARUS NO.	QUANTITY	MOTOR HP	BASE	ADJUSTED	TOTAL HP	(BASE CASE)
01 FEEDSTOCK HANDLING 100 UNIT								
SCALE-ADJUSTMENT FACTOR 1.24, EXPONENT 0.80, OVERALL COST MULTIPLIER 1.19								
J-101	CHIP CONVEYOR	CO-101	6	62	$226.8	$289.1	373.31	216 TPH
J-102	BELT CONVEYOR	CO-102	1	6	$49.0	$62.5	6.22	60 TPH
F-103	VIB SCREEN	VS-103	3	31	$75.3	$96.0	93.33	
G-108	TRUCK SCALE	S-108	1	0	$44.4	$52.9	0.00	60 TONS
G-109	RAIL CAR SCALE	S-109	1	0	$42.2	$50.3	0.00	60 TONS
P-101	FLUME PUMP	CP-101	1	25	$6.8	$8.1	24.89	2000 GPM @ 15 FT
G-104ABS	FRONT END LOADER	GM-104	3	0	$294.9	$351.3	0.00	
G-105	MAG CHIP CLNR	GS-105	1	0	$9.4	$11.2	0.00	
M-106AB	YARD SWITCH ENG	GM-106	2	0	$1,282.2	$1,527.3	0.00	
G-107	WOOD CHIP TRUCK UNLDR	GY-107	1	0	$8.5	$10.2	0.00	
PURCHASED EQUIPMENT COST				***498***	***$2,039.6***	***$2,458.7***	***497.75***	
DILUTE ACID PRETREATMENT BLOCK								
01 ENZ HYDR-DIL ACID PTRMT 100 UNIT								
SCALE-ADJUSTMENT FACTOR 1.24, EXPONENT 0.80, OVERALL COST MULTIPLIER 1.19								
H-201	1ST STG WTR PRHTR	HE-101	1		$12.7	$16.2	0.00	840 SF
H-202	2ND STG WTR PRHTR	HE-102	1		$24.3	$31.0	0.00	2100 SF
T-202	FLASH VSL C=124600	VT-102	1		$106.5	$126.9	0.00	6100 GAL
T-203	FLASH VSL	VT-103	1		$80.1	$95.4	0.00	3800 GAL
T-204	FLASH VSL	VT-104	1		$34.5	$41.1	0.00	12900 GAL
P-201	WATER FEED PMP	CP-101	1	187	$11.7	$14.9	186.66	750 GPM @ 575 FT
P-201S	WATER FEED PMP (SPARE)	CP-101	1	187	$11.7	$14.9	0.00	750 GPM @ 575 FT
PURCHASED EQUIPMENT COST				***187***	***$281.5***	***$340.4***	***186.66***	
02 ENZ HYDR-DIL ACID PTRMT 100 UNIT								
SCALE-ADJUSTMENT FACTOR 1.24, EXPONENT 0.80, OVERALL COST MULTIPLIER 1.19								
T-201	CHIP BIN	VT-101	1		$72.8	$92.8	0.00	11000 GAL
J-203	SCREW FEEDER C=0	CO-103	3	6	$0.0	$0.0	18.67	123 TPH
J-204	SCREW DIGSTR C =0	CO-104	3	37	$0.0	$0.0	111.99	494 TPH
C-201	KNIFE MILL (DISK RFNR) C=4050	CR-101	3	498	$1,038.6	$1,237.1	1493.25	
R-201	PREHYD REACTOR C=18500	AT-101	3	2	$47.4	$56.5	5.60	180 GAL
PURCHASED EQUIPMENT COST				***1630***	***$1,158.9***	***$1,386.4***	***1629.51***	
03 ENZ HYDR-DIL ACID PTRMT 200 UNIT								
SCALE-ADJUSTMENT FACTOR 1.24, EXPONENT 0.80, OVERALL COST MULTIPLIER 1.19								
T-221	LIME TANK	AT-201	1	50	$89.8	$114.5	49.78	11000 GAL
T-222	NEUTRALIZATION TANK	AT-202	1	37	$74.6	$95.1	37.33	7000 GAL
P-222	LIME PUMP	CP-202	1	2	$2.7	$3.4	2.49	50 GPM @ 100 FT.
P-222S	LIME PUMP (SPARE)	CP-202	1	2	$2.7	$3.4	0.00	50 GPM @ 100 FT.
P-223	THKNR PUMP	CP-203	1	31	$4.4	$5.6	31.11	700 GPM @ 100 FT.
P-223S	THKNR PUMP (SPARE)	CP-203	1	31	$4.4	$5.6	0.00	700 GPM @ 100 FT.
T-221	THICKENER	T -201	1		$52.1	$66.4	0.00	30 FT DIA.
C-222	SOLID BOWL CENTRIFUGE	CT-202	1	37	$85.2	$108.6	37.33	
P-224	OVERFLOW PUMP	CP-204	1	4	$2.7	$3.4	3.73	30 GPM @ 100 FT.
P-224S	OVERFLOW PUMP (SPARE)	CP-204	1	4	$2.7	$3.4	0.00	30 GPM @ 100 FT.
T-222	SUGAR RECEIVER	VT-202	1		$203.5	$259.4	0.00	325000 GAL
P-225	SUGAR TRANS PUMP	CP-205	1	31	$4.5	$5.7	31.11	725 GPM @ 100 FT.
P-225S	SUGAR TRANS PUMP (SPARE)	CP-205	1	31	$4.5	$5.7	0.00	725 GPM @ 100 FT.
PURCHASED EQUIPMENT COST				***193***	***$533.8***	***$680.3***	***192.88***	
04 ENZ HYDR-DIL ACID PTRMT 200 UNIT								
SCALE-ADJUSTMENT FACTOR 1.24, EXPONENT 0.80, OVERALL COST MULTIPLIER 1.19								
C-221	SOLID BOWL CNTRFG C=455500	CT-201	2	373	$778.7	$927.6	746.63	
T-221	PREHYD BIN C=143200	VT-201	1		$122.4	$145.8	0.00	90000.00 GALLONS
P-221AB	OFLW PP	CP-201	2	19	$11.6	$14.8	37.33	310 GPM @ 100 FT.
P-221ABS	OFLW PP (SPARE)	CP-201	2	19	$11.6	$14.8	0.00	310 GPM @ 100 FT.
F-221	POLISHING FILTER	F-201	1	4	$50.4	$64.2	3.73	100 SF
E-221	THICKENER HX	HE-201	1		$55.2	$70.4	0.00	1175 SF
E-222	FILTER COOLER	HE-202	1		$43.3	$55.2	0.00	880 SF
J-221	SCREW CONVEYOR	CO-201	1	249	$192.1	$244.8	248.88	494 TPH
J-222	SCREW CONVEYOR	CO-202	1	37	$51.9	$66.1	37.33	69.00 TPH
	ROTATING EQPT SPARE PARTS				$5.3	$6.3	0.00	
PURCHASED EQUIPMENT COST				***1074***	***$1,322.6***	***$1,610.0***	***1073.90***	
ENZYME PRODUCTION BLOCK								
01 ENZYME PRODUCTION SYSTEM 300 UNIT								
SCALE-ADJUSTMENT FACTOR 0.27, EXPONENT 0.60, OVERALL COST MULTIPLIER 0.45								
T-314	ENZ PROD RCTRS	AT -714	40	40	$11,593.4	$5,270.4	1612.62	114000 GAL (50 PSIG)
T-343	NH3 STRG TANK	HT-743			$20.7	$9.4	0.00	16500 GAL
P-344A	NH3 MTRG PUMP	CP-744	1	1	$2.6	$1.2	0.54	5 GPM @ 75 FT.
P-344S	NH3 MTRG PUMP (SPARE)	CP-744	1	1	$2.6	$1.2	0.00	5 GPM @ 75 FT.

SIMULTANEOUS SACCHARIFICATION AND FERMENTATION, DILUTE ACID PRETREATMENT

CAPITAL EQUIPMENT LIST			BASE		EQUIPMENT COST ($k)		ADJUSTED	DESIGN DATA
P&ID NO.	ITEM	ICARUS NO.	QUANTITY	MOTOR HP	BASE	ADJUSTED	TOTAL HP	(BASE CASE)
T-344	NUTRIENT STRG TNK	VT -744	1		$44.9	$20.4	0.00	19800 GAL
P-345A	NUTRIENT PMP	CP -745	1	1	$2.6	$1.2	0.54	15 GPM @ 75 FT.
P-345S	NUTRIENT PMP (SPARE)	CP -745	1	1	$2.6	$1.2	0.00	15 GPM @ 75 FT.
T-305	MIX TANK	VT -705			$25.0	$11.4	0.00	25000 GALLONS
P-307A	METERING PUMP	CP -707	1	1	$2.7	$1.2	1.34	120 GPM @ 75 FT.
P-307S	METERING PUMP (SPARE)	CP -707	1	1	$2.7	$1.2	0.00	120 GPM @ 75 FT.
T-312	SURGE TANK	VT-712			$61.7	$28.0	0.00	25000 GAL
P-313A	NUTRNT FEED PMP	CP -713	1	0	$2.6	$1.2	0.36	120 GPM @ 75 FT.
P-313S	NUTRNT FEED PMP (SPARE)	CP -713	1	1	$2.6	$1.2	0.00	120 GPM @ 75 FT.
U-308	PASTEURIZER C=75000	HE-708	1		$64.1	$29.1	0.00	1000 SF (150 PSIG)
T-326	1STST SDTNK C=60000	AT -726	1	0	$51.3	$23.3	0.27	15 GALLONS
T-327	2NDST SDTNK C=240000	AT -727	1	1	$23.5	$10.7	1.34	275 GALLONS
T-328	3RDST SDTNK C=760000	AT -728	1	5	$68.8	$31.3	5.38	6000 GAL
P-320A-AN	RECYCLE PUMPS	CP -720	40	5	$248.0	$112.7	215.02	500 GPM @ 100 FT.
P-320AS-ANS	RECYCLE PUMPS (SPARE)	CP -720	40	5	$248.0	$112.7	0.00	500 GPM @ 100 FT.
H-300A-AN	FERMENTER COOLERS	HE -700	40		$520.0	$236.4	0.00	256 SF (150 PSIG)
PURCHASED EQUIPMENT COST				***1837***	***$12,990.4***	***$5,905.4***	***1837.40***	
02 AIR COMPRESSION SYSTEM 300 UNIT								
SCALE-ADJUSTMENT FACTOR 0.27, EXPONENT 0.60, OVERALL COST MULTIPLIER 0.45								
P-321AB	AIR COMPRESSOR	GC -301	2	2150	$4,788.6	$2,176.9	4300.32	57000.00 CFM @ 35 PSIG
T-321AB	INTRCLRS	HE -301	4		$56.4	$25.6	0.00	1000 SF
PURCHASED EQUIPMENT COST				***4300***	***$4,845.0***	***$2,202.5***	***4300.32***	
03 CHILLED WATER REFRIG SYSTEM								
SCALE-ADJUSTMENT FACTOR 0.27, EXPONENT 0.60, OVERALL COST MULTIPLIER 0.45								
T-330	CHLD WTR C=150000	VT -100	8		$1,025.8	$466.3	0.00	17600 GAL
P-331	CHLDWTR CIRCPMP	CP -101	2	34	$26.8	$12.2	67.19	2000 GPM @ 200 FT.
P-331S	CHLDWTR CIRCPMP (SPARE)	CP -101	1	34	$13.4	$6.1	0.00	2000 GPM @ 200 FT.
PURCHASED EQUIPMENT COST				***67***	***$1,066.0***	***$484.6***	***67.19***	
04 HYDROLYSIS/FERMENTATION 400 UNIT								
SCALE-ADJUSTMENT FACTOR 1.09, EXPONENT 0.60, OVERALL COST MULTIPLIER 1.06								
T-413A-P	SSF REACTORS	VT -413	29		$11,405.6	$12,040.6	0.00	474000.0 GALLON5
T-405AB	SURGE BINS	VT -405	2		$150.4	$158.8	0.00	105000 GALLONS
J-405	CLOSED-BLT FEED CONVEYORS	CO -406	2	3	$73.4	$77.5	6.57	60 TPH
J-406A-P	REACTOR FEED CONVEYORS	CO -406	29	8	$73.6	$77.7	238.05	54 TPH
M-414	AGITATORS C=57000	MOT-414	87	109	$2,338.8	$2,469.0	9522.08	
H-401A-P	REACTOR COOLERS	HE -400	29		$435.2	$459.4	0.00	760 SF
P-430A-P	REACTOR CIRC PUMPS	CP -430	29	44	$91.2	$96.3	1269.61	1000 GPM @ 100 FT.
P-430AS-PS	REACTOR CIRC PUMPS (SPARE)	CP -430	29	44	$91.2	$96.3	0.00	1000 GPM @ 100 FT.
C-420	HYD CENTRIFUGE	CT -420	2		$477.2	$503.8	0.00	54 IN DIA.
P-424A	HYD PUMP	CP -424	1	22	$4.2	$4.4	21.89	400 GPM @ 125 FT.
P-424S	HYD PUMP (SPARE)	CP -424	1	22	$4.2	$4.4	0.00	400 GPM @ 125 FT.
J-425	CLOSED-BLT LIGNIN CONVEYOR	CO -425		2	$31.6	$33.4	0.00	60 TPH
T-415	PLAST TANK ACID TANK	VT -415			$7.0	$7.4	0.00	5000 GAL
P-417A	ACID TRANS PMP	CP -417	1	5	$2.9	$3.1	5.47	100 GPM @ 100 FT
P-417S	ACID TRANS PMP (SPARE)	CP -417	1	5	$2.9	$3.1	0.00	100 GPM @ 100 FT
T-416	CAUSTIC TANK	VT -416	1		$7.0	$7.4	0.00	5000 GALLONS
P-418A	CAUST TRANS PMP	CP -418	1	5	$3.0	$3.2	5.47	100 GPM @ 100 FT
P-418AS	CAUST TRANS PMP (SPARE)	CP -418	1	5	$3.0	$3.2	0.00	100 GPM @ 100 FT
P-419	HYD TRANS PUMP	CP -419	3	137	$40.2	$42.4	410.43	2000 GPM @ 200 FT.
P-419S	HYD TRANS PUMP (SPARE)	CP -419	1	137	$13.4	$14.1	0.00	2000 GPM @ 200 FT.
T-427	PAST WATER STRG TANK	VT -427	1		$4.0	$4.2	0.00	1500 GAL
R-423	PASTEURIZER C=20000	HE -423	1		$17.1	$18.0	0.00	1000 SF
	ROTATING EQPT SPARE PARTS				$556.5	$587.5		
PURCHASED EQUIPMENT COST				***11480***	***$15,833.7***	***$16,715.1***	***11479.58***	
06 HYDROLYSIS/FERMENTATION 400 UNIT								
SCALE-ADJUSTMENT FACTOR 1.09, EXPONENT 0.60, OVERALL COST MULTIPLIER 1.06								
T-504	BEER WELL (CONE ROOF)	VT-504	1		$64.7	$68.3	0.00	225000 GAL
H-501AH	HX U-TUBE	HE-501	16		$337.6	$356.4	0.00	1400 FT2
T-502AH	FERMENTERS	VT-502	16		$517.6	$546.4	0.00	225000 GAL
T-505	STERILIZATION TANK	VT-505	1		$19.8	$20.9	0.00	10000 GAL
T-506	CLEANING TANK	VT-506	1		$19.8	$20.9	0.00	10000 GAL
P-504A	BEERWELL PUMP	CP-504	1	27	$6.3	$6.7	27.36	1500 GPM @ 50 FT.
P-504S	BEERWELL PUMP (SPARE)	CP-504	1	27	$6.3	$6.7	27.36	1500 GPM @ 50 FT.
P-501AH	HX PUMP	CP-501	16	22	$32.8	$34.6	350.24	670 GPM @ 80 FT.
P-501AH/S	HX PUMP (SPARE)	CP-501	16	22	$32.8	$34.6	0.00	670 GPM @ 80 FT.
P-505A	DIST FEED PUMP	CP-505	1	219	$25.5	$26.9	218.90	1500 GPM @ 400 FT.
P-505S	DIST FEED PUMP (SPARE)	CP-505	1	219	$25.5	$26.9	0.00	1500 GPM @ 400 FT.
P-506A	STERILIZATION PUMP	CP-506	1	16	$2.5	$2.6	16.42	200 GPM @ 200 FT.
P-506S	STERILIZATION PUMP (SPARE)	CP-506	1	16	$2.5	$2.6	0.00	200 GPM @ 200 FT.

SIMULTANEOUS SACCHARIFICATION AND FERMENTATION, DILUTE ACID PRETREATMENT

CAPITAL EQUIPMENT LIST

P&ID NO.	ITEM	ICARUS NO.	BASE QUANTITY	MOTOR HP	EQUIPMENT COST ($k) BASE	EQUIPMENT COST ($k) ADJUSTED	ADJUSTED TOTAL HP	DESIGN DATA (BASE CASE)
P-503A	CLEANING PUMP	CP-503	1	8	$3.1	$3.3	8.21	300 GPM @ 50 FT
P-503S	CLEANING PUMP (SPARE)	CP-503	1	8	$3.1	$3.3	0.00	300 GPM @ 50 FT
P-507A	SCRBRCYC PUMP	CP-507	1	3	$2.5	$2.6	3.28	150 GPM @ 50 FT
P-507S	SCRBRCYC PUMP (SPARE)	CP-507	1	3	$2.5	$2.6	0.00	150 GPM @ 50 FT
F-503A-Q	STRAINER	CP-503	1	1	$58.1	$61.4	1.09	20 GPM @ 75 FT
PURCHASED EQUIPMENT COST				***653***	***$1,163.0***	***$1,227.8***	***652.86***	
07 HYDROLYSIS/FERMENTATION 400 UNIT								
SCALE-ADJUSTMENT FACTOR 1.09, EXPONENT 0.60, OVERALL COST MULTIPLIER 1.06								
T-501	C02 OFFGAS SCRBR	TK-501			$22.0	$23.2	0.00	5 FT. DIA. X 25 FT HT.
T-500	YEAST MIX TANK	AT-501	1	44	$58.6	$61.9	43.78	10000 GAL
PURCHASED EQUIPMENT COST				***44***	***$80.6***	***$85.1***	***43.78***	
XYLOSE FERMENTATION 500 UNIT								
SCALE-ADJUSTMENT FACTOR 0.51, EXPONENT 0.60, OVERALL COST MULTIPLIER 0.66								
T-550	YEAST MIX TANK	AT-501	1	20	$58.6	$39.0	20.26	10000 GAL
H-551AH	HX U-TUBE	HE-501	14		$337.6	$224.4	0.00	1400 FT2
T-502AH	FERMENTERS	VT-502	14		$1,508.8	$1,003.1	0.00	225000 GAL
P-554A	ETOH TRANSFER PUMP	CP-504	1	0	$6.3	$4.2	0.00	1500 GPM @ 50 FT.
P-554S	ETOH TRANSFER PUMP (SPARE)	CP-504	1	0	$6.3	$4.2	0.00	1500 GPM @ 50 FT.
P-551AH	HX PUMP	CP-501	14	0	$32.8	$21.8	0.00	670 GPM @ 80 FT.
P-551AH/S	HX PUMP (SPARE)	CP-501	14	0	$32.8	$21.8	0.00	670 GPM @ 80 FT.
PURCHASED EQUIPMENT COST				***20***	***$1,983.2***	***$1,318.5***	***20.26***	
08 ETHANOL PURIFICATION 600 UNIT								
SCALE-ADJUSTMENT FACTOR 0.89, EXPONENT 0.60, OVERALL COST MULTIPLIER 0.93								
R-601	BEER COLUMN (SIEVE TRAY)	TW-601			$151.4	$140.9	0.00	16 TRAYS;12 FT.DIA. @ 50
R-602	RECT COLUMN (SIEVE TRAY)	TW-602			$142.8	$132.9	0.00	24 TRAYS; 10 FT. DIA. @ 6
T-602	BEERCOL REFLUX DRUM	HT-602			$4.4	$4.1	0.00	800 GAL
T-603	FUSLOIL DECANTR	HT-603			$4.4	$4.1	0.00	870 GAL
T-604	MOLSIEVE FEEDTANK	VT-604			$21.6	$20.1	0.00	38000 GAL
T-605	RECTCOL REFDRUM	HT-605			$10.3	$9.6	0.00	3800 GALLONS; 6 FT.D.X 1
P-601A	BEERCOL BTMSPMP	CP-601	1	53	$7.3	$6.8	53.21	1450.00 GPM @115 FT H.
P-601S	BCOL BTMSPMP SPARE	CP-601	1	53	$7.3	$6.8	0.00	1450.00 GPM @115 FT H.
P-603	BEERCOL REFPUMP	CP-603	1	13	$3.8	$3.5	13.30	220 GPM @136 FT
P-603S	BEERCOL REFPUMP (SPARE)	CP-603	1	13	$3.8	$3.5	0.00	220 GPM @136 FT
P-604A	WASHRET PUMP	CP-604	1	4	$3.2	$3.0	4.43	30 GPM @ 136 FT.
P-604S	WASHRET PUMP (SPARE)	CP-604	1	4	$3.2	$3.0	0.00	30 GPM @ 136 FT.
P-605A	FUSLOIL PUMP	CP-605	1	3	$2.8	$2.6	2.66	5 GPM @ 90 FT.
P-605AS	FUSLOIL PUMP (SPARE)	CP-605	1	3	$2.8	$2.6	0.00	5 GPM @ 90 FT.
P-606A	MOLSIVE PUMP	CP-606	1	7	$2.2	$2.0	6.65	65 GPM @ 260 FT.
P-606S	MOLSIVE PUMP (SPARE)	CP-606	1	7	$2.2	$2.0	0.00	65 GPM @ 260 FT.
P-607A	RCT COL BTM PMP	CP-607A	1	4	$3.0	$2.8	4.43	120 GPM @90 FT.
P-607S	RCT COL BTM PMP (SPARE)	CP-607A	1	4	$3.0	$2.8	0.00	120 GPM @90 FT.
P-608A	RCTCOLREFPMP	CP-608	1	13	$3.9	$3.6	13.30	260 GPM @ 136 FT.
P-608S	RCTCOLREFPMP (SPARE)	CP-608	1	13	$3.9	$3.6	0.00	260 GPM @ 136 FT.
H-605	BEER COL CONDENSER	HE-605			$60.4	$56.2	0.00	6000 SF
H-609	RECT COL REBOILER	HE-609			$95.8	$89.1	0.00	4550 SF
G-601	AIR DRYER MOL SIEVES	{G GZ-601			$2,137.1	$1,988.5	0.00	50000.00 CFM
PURCHASED EQUIPMENT COST				***98***	***$2,680.6***	***$2,494.3***	***98.00***	
09 ETHANOL PURIFICATION 600 UNIT								
SCALE-ADJUSTMENT FACTOR 0.89, EXPONENT 0.60, OVERALL COST MULTIPLIER 0.93								
T-601	DEGASSER DRUM	VT-601			$10.5	$9.8	0.00	3800 GAL
H-602	DEG DRM CONDENSER	HE-602			$18.2	$16.9	0.00	1500 SF
H-603	REBOILERS	HE-603			$208.4	$193.9	0.00	2 X 5000 SF
H-606	BEER COL VENT CNDSR	HE-606			$77.1	$71.7	0.00	2 X 3500 SF
H-607	FUSLOIL COOLER	HE-607			$8.2	$7.6	0.00	300 SF
H-610	RECT COL CONDENSER	HE-610			$38.5	$35.8	0.00	3500 SF
H-611	RECT COL VENT CNDNSR	HE-611			$77.1	$71.7	0.00	2 X 3500 SF
H-613	FEED PREHEATER	HE-613			$632.1	$588.2	0.00	4 X 5000 SF
PURCHASED EQUIPMENT COST				***0***	***$1,070.1***	***$995.7***	***0.00***	
12 TANKAGE 700 UNIT								
SCALE-ADJUSTMENT FACTOR 1.24, EXPONENT 0.50, OVERALL COST MULTIPLIER 1.12								
T-701AS	PRD STRTK	VT-701	2		$251.2	$280.2		510000 GALLONS
T-708	SLOPS TANK	VT-708			$45.8	$51.1		127000 GAL
T-707	FIRE WATER TANK	VT-707			$125.6	$140.1		510000 GALLONS
T-703	H2S04 STRGTNK	VT-703			$28.3	$31.6		62000.00 GALLONS
T-705	FUSELOILSTRG	VT-705			$22.8	$25.4		42000 GAL
T-709	DIESEL FUEL TANK	VT-709			$15.6	$17.4		21000.00 GALLONS
T-710	GASOLINE STRGTANK	VT-710			$55.2	$61.6		169200.0 GALLONS
P-701AB	ETOHPROD	CP-701	2	19	$7.6	$8.5	37.33	525 GPM @ 70 FT.

SIMULTANEOUS SACCHARIFICATION AND FERMENTATION, DILUTE ACID PRETREATMENT

CAPITAL EQUIPMENT LIST			BASE		EQUIPMENT COST ($k)		ADJUSTED	DESIGN DATA
P&ID NO.	ITEM	ICARUS NO.	QUANTITY	MOTOR HP	BASE	ADJUSTED	TOTAL HP	(BASE CASE)
P-701S	ETOHPROD (SPARE)	CP-701	1	19	$3.8	$4.2	0.00	525 GPM @ 70 FT.
P-708A	SLOPS TRAN PMP	CP-708	1	6	$3.0	$3.3	6.22	100 GPM @ 90 FT.
P-708S	SLOPS TRAN PMP (SPARE)	CP-708	1	6	$3.0	$3.3	0.00	100 GPM @ 90 FT.
P-707A	FIREWATER PMP	CP-707	1	62	$5.5	$6.1	62.22	600 GPM @ 227 FT.
P-707AS	FIREWATER PMP (SPARE)	CP-707	1	62	$5.5	$6.1	0.00	600 GPM @ 227 FT.
P-703A	H2SO4 TRANS	CP-703	1	6	$3.0	$3.3	6.22	100 GPM @ 90 FT.
P-703S	H2SO4 TRANS (SPARE)	CP-703	1	6	$3.0	$3.3	0.00	100 GPM @ 90 FT.
P-705A	FUSLOIL EXPT	CP-705	1	4	$2.7	$3.0	3.73	100 GPM @ 70 FT.
P-705S	FUSLOIL EXPT (SPARE)	CP-705	1	4	$2.7	$3.0	0.00	100 GPM @ 70 FT.
P-709A	DIESEL FUEL	CP-709	1	4	$3.0	$3.3	3.73	25 GPM @ 115 FT.
P-709S	DIESEL FUEL (SPARE)	CP-709	1	4	$3.0	$3.3	0.00	25 GPM @ 115 FT.
P-710A	GASOL BLNDNG	CP-710	1	2	$2.6	$2.9	2.49	30 GPM @ 70 FT.
P-710S	GASOL BLNDNG (SPARE)	CP-710	1	2	$2.6	$2.9	0.00	30 GPM @ 70 FT.
F-703	AIR DRYER DESC AIR FILTER	AD-703			$35.1	$39.2		1500 CFM
PURCHASED EQUIPMENT COST				*122*	*$630.6*	*$703.4*	*121.95*	
13 WASTE TREATMENT 800 UNIT								
SCALE-ADJUSTMENT FACTOR 0.86, EXPONENT 0.60, OVERALL COST MULTIPLIER 0.91								
T-801	HOTWELL DECANTER	VT-801	1	0	$98.5	$90.1	0.00	27200 GAL
T-802	MIXED SOLVENT TANK	VT-802	1	0	$52.9	$48.4	0.00	169200 GAL
T-807	BIOTREATER	VT-807	1	0	$99.3	$90.8	0.00	350000 GAL
T-809	FILTER FEED TANK	VT-808	1	0	$27.8	$25.4	0.00	30000 GAL
P-808A	REACTOR FEED PUMP	CP-808	1		$2.0	$1.8	0.00	
P-808S	REACTOR FEED PUMP (SPARE)	CP-808	1		$2.0	$1.8	0.00	
P-809A	REACTOR RECYCLE PUMP	CP-809	1		$2.0	$1.8	0.00	
P-809AS	REACTOR RECYCLE PUMP (SPAR	CP-809	1		$2.0	$1.8	0.00	
P-813A	SLUDGE PUMP	CP-813	1	13	$4.0	$3.7	12.92	650 GPM @ 45 FT.
P-813AS	SLUDGE PUMP (SPARE)	CP-813	1	13	$4.0	$3.7	0.00	650 GPM @ 45 FT.
P-816A	FINAL EFF PUMP	CP-816	1	17	$3.9	$3.6	17.22	700 GPM @ 90 FT
P-816S	FINAL EFF PUMP (SPARE)	CP-816	1	17	$3.9	$3.6	0.00	700 GPM @ 90 FT
P-817A	LP VENT BLWR	FN-817	1	65	$8.1	$7.4	64.59	2400 CFM
P-817S	LP VENT BLWR (SPARE)	FN-817	1	65	$8.1	$7.4	0.00	2400 CFM
H-805	STILLAGE COOLER	HE-805	1	0	$315.8	$288.7	0.00	5000 FT2
H-806	SOLVENT COOLER	HE-806	1	0	$6.5	$5.9	0.00	40 FT2
F-804	ROTARY FILTER	F-804	1	3	$38.3	$35.0	2.58	10 FT2
T-808	SEC CLARIFIER	T-808	1	0	$101.3	$92.6	0.00	80 FT DIA.
M-808	AGITATOR	AG-812	1	4	$2.3	$2.1	4.31	
F-810	BELT FILTER		1		38.3	$35.0	0.00	2 METERS
P-810A	FILTER FEED PUMP		1	25	$5.0	$4.6	25.00	85 GPM @ 30 FT.
P-810AS	FILTER FEED PUMP (SPARE)		1	25	$5.0	$4.6	0.00	85 GPM @ 30 FT.
T-803	EQUAL TANK	VT-803	1	0	$174.9	$159.9	0.00	600000 GAL
T-804	ANAEROBIC REACTR	VT-804	1	0	$303.2	$277.2	0.00	160000 GAL
T-805	SURGE DRUM	VT-805	1	0	$18.3	$16.7	0.00	10000 GAL
T-806	KO POT OFFGAS	VT-806	1	0	$7.6	$6.9	0.00	1500 GAL
PURCHASED EQUIPMENT COST				*127*	*$1,335.0*	*$1,220.5*	*126.62*	
14 WASTE TREATMENT 800 UNIT								
SCALE-ADJUSTMENT FACTOR 0.86, EXPONENT 0.60, OVERALL COST MULTIPLIER 0.91								
T-810	LP VENT KO DRUM	VT-810	1	0	$4.6	$4.2	0.00	475 GAL
G-804	2-STAGE VACUUM SYSTEM	EJ-804	1	0	$8.0	$7.3	0.00	115.00 LB/H
H-801,802	EVAP 1,2-EFF HTRS	HE-801	2	0	$514.8	$470.6	0.00	4 X 6500 SF
H-803	3RD EFF HTR	HE-803	1	0	$813.1	$743.4	0.00	4 X 6500 SF
H-804	EVAP CONDNSR	HE-804	1	0	$873.5	$798.6	0.00	5 X 5600 SF
T-801	1ST EFF EVAP	VT-801	1	0	$34.3	$31.4	0.00	20300.00 GALLONS
T-802	2ND EFF EVAP	VT-802	1	0	$58.0	$53.0	0.00	34400 GALLONS
T-803	3RD EFF EVAP	VT-803	1	0	$90.8	$83.0	0.00	56400.00 GALLONS
P-801A	1ST EFFECT PUMP	CP-801	1	17	$5.2	$4.8	17.22	1050 GPM @ 45 FT.
P-801S	1ST EFFECT PUMP (SPARE)	CP-801	1	17	$5.2	$4.8	0.00	1050 GPM @ 45 FT.
P-802A	2ND EFFECT PUMP	CP-802	1	13	$4.0	$3.7	12.92	600 GPM @ 70 FT.
P-802S	2ND EFFECT PUMP (SPARE)	CP-802	1	13	$4.0	$3.7	0.00	600 GPM @ 70 FT.
P-803A	3RD EFFECT PUMP	CP-803	1	6	$2.8	$2.6	6.46	150 GPM @ 125 FT.
P-803S	3RD EFFECT PUMP (SPARE)	CP-803	1	6	$2.8	$2.6	0.00	150 GPM @ 125 FT.
P-805A	EVAP WASTE PUMP	CP-805	1	34	$6.0	$5.5	34.45	1250 GPM @ 90 FT.
P-805S	EVAP WASTE PUMP (SPARE)	CP-805	1	34	$6.0	$5.5	0.00	1250 GPM @ 90 FT.
P-806A	SOLVENT PUMP	CP-806	1	1	$2.6	$2.4	0.86	5 GPM @ 80 FT.
P-806S	SOLVENT PUMP (SPARE)	CP-806	1	1	$2.6	$2.4	0.00	5 GPM @ 80 FT.
P-807A	MIXED SOLVENT PUMP	CP-807	1	9	$3.5	$3.2	8.61	300 GPM @ 90 FT.
P-807S	MIXED SOLVENT PUMP (SPARE)	CP-807	1	9	$3.5	$3.2	0.00	300 GPM @ 90 FT.
PURCHASED EQUIPMENT COST				*81*	*$2,445.3*	*$2,235.6*	*80.52*	
15 UTILITIES 900 UNIT								
SCALE-ADJUSTMENT FACTOR 0.89, EXPONENT 0.80, OVERALL COST MULTIPLIER 0.91								
T-901	PROCESS WATER TANK	VT-901			$137.4	$125.1		585000 GALLONS

SIMULTANEOUS SACCHARIFICATION AND FERMENTATION, DILUTE ACID PRETREATMENT

CAPITAL EQUIPMENT LIST

P&ID NO.	ITEM	ICARUS NO.	BASE QUANTITY	MOTOR HP	EQUIPMENT COST ($k) BASE	EQUIPMENT COST ($k) ADJUSTED	ADJUSTED TOTAL HP	DESIGN DATA (BASE CASE)
T-902	BLWDN FLASH DRUM	VT-902			$5.6	$5.1		530 GALLONS
T-903	HYDRAZINE DRUM	VT-903			$7.2	$6.6		250 GAL
T-904	CONDSURGE DRUM	HT-904			$20.8	$18.9		18000 GAL
T-905	BACKWASH TRANS TNK	VT-905			$10.2	$9.3		12000 GAL
T-906,907	AIR RECEIVERS	VT-906	2		$24.4	$22.2		2400 GAL
P-902A	PROC WATR PMP	CP-902	1	53	$5.9	$5.4	53.34	2400 GPM @ 70 FT.
P-902S	PROC WATR PMP (SPARE)	CP-902	1	53	$5.9	$5.4	0.00	2400 GPM @ 70 FT.
P-903A	PROC WAT CIRC PMP	CP-903	1	222	$27.1	$24.7	222.24	5000 GPM @ 136 FT.
P-903S	PROC WAT CIRC PMP (SPARE)	CP-903	1	222	$27.1	$24.7	0.00	5000 GPM @ 136 FT.
P-904AB	BACKWASH FEED PMP	CP-904	2	89	$21.2	$19.3	177.79	6000 GPM @ 48 FT.
P-905AB	BACKWASH TRN PMP	CP-905	2	1	$4.0	$3.6	1.78	50 GPM @ 48 FT.
P-906A	BLOWDOWN PMP	CF-906	1	1	$2.1	$1.9	1.33	50 GPM @ 57 FT.
P-906S	BLOWDOWN PMP (SPARE)	CF-906	1	1	$2.1	$1.9	0.00	50 GPM @ 57 FT.
P-907	HYDRAZINE TRANS PMP	CP 907	1	1	$3.7	$3.4	1.33	5 GPM @ 45 FT.
P-908A	HP BFW PUMP	CP -908	1	1333	$103.0	$93.7	1333.43	1700 GPM @ 2838 FT.
P-908S	HP BFW PUMP (SPARE)	CP -908	1	1333	$103.0	$93.7	0.00	1700 GPM @ 2838 FT.
P-909A	DEAR FD PUMP	CP -909	1	22	$4.9	$4.5	22.22	1700 GPM @ 45 FT.
P-909S	DEAR FD PUMP (SPARE)	CP -909	1	22	$4.9	$4.5	0.00	1700 GPM @ 45 FT.
P-910A	TURB COND PMP	CF -910	1	18	$3.9	$3.5	17.78	650 GPM @ 90 FT
P-910AS	TURB COND PMP (SPARE)	CF -910	1	18	$3.9	$3.5	0.00	650 GPM @ 90 FT
PC911	AIR COMPRESSOR (RECIP)	GC -911	1	178	$203.8	$185.5	177.79	1000 CFM, 125 PSIG DISC
P-912A-H	CW CIRC	CF -912	8	311	$279.2	$254.1	2489.07	7500 GPM @ 136 FT.
P-913AB	WELL WATR PUMP	CP -913	2	36	$12.0	$10.9	71.12	2000 GPM @57 FT.
P-913ABS	WELL WATR PUMP (SPARE)	CP -913	1	36	$6.0	$5.5	0.00	2000 GPM @57 FT.
F-901	SAND FILTER	VT -901	1		$36.3	$33.0	0.00	54000 GAL
T-905	COND COLL TNK	VT -905			$8.8	$8.0	0.00	2000 GAL
	LINING MAT'L	LIN-905			$38.4	$34.9	0.00	2000 SF
G-905AB	PUMPS	CP -905	2	36	$19.2	$17.5	71.12	1700 GPM @ 70 FT.
G-907,908,909	TANKS	VT -907	3		$22.5	$20.5	0.00	150 GAL
G-907,908,909	DIAPHRAGM PUMPS	P -907	6	0	$10.8	$9.8	0.69	5 GPM
G-910	INST AIR DRYER	AD -910			$21.0	$19.1		600 CFM
GU-903AB $425000 EACH PERMUTIT								$425.0
COOLING TOWER MATERIALS (MTL-$400000 L-S332000 FLD ERECT)								$400.0
COOLING TOWER LABOR (MTL-$400000 L-S332000 FLD ERECT)								$332.0
GZ-911 TURBOGENERATOR $4800000 ALL SHOP-FAB PARTS								$4,800.0
HB-901AB STEAM BOILERS FIELD ERECTED								
FOUNDATION CONDENSER ADD $800000								$800.0
BOILER MATERIALS (MTL-$5200000 L-11050000)			2					$5,200.0
BOILER LABOR (MTL-$5200000 L-$1050000)			2					$1,050.0
FLAKT SYSTEM FABRICATED PIECES $2400000 (INSTALLED AS MULTIPLE SPRAY DRYERS BY SYSTEM)								$2,400.0
BOILERS AND AUXILIARIES INSTALLED AND COSTED BY ICARUS SYSTEM (BADGER COSTS FOR REFERENCE ONLY)								
G-911	TURBO-GEN	EG-911			$5,412.3	$4,925.9	0.00	44500 KVA
G-912	COOLING TOWER SYST (FANS)	CTW-912	10	111	$1,067.6	$971.7	1111.19	60000 GPM (30 DEG.F, 74
	COOLING TOWER SYST (PUMPS)	CTW-912	3	356			1066.75	
B-901AB	STM BOILER	STB-901	2		$3,072.0	$2,795.9		400000.0 LB/H (1000 PSIG
G-903AB	DENINERALIZER C=850000	WTS-903			$726.6	$661.3		15000.00 GPH
T-906	DEAERATOR C=120000	VT-903			$102.6	$93.4		17000 GAL @ 650 DEG.F
D-901	SPRAY DRYER C=400000	D-901	6		$2,051.6	$1,867.2		EVAP.RATE 3500 LB/H
	ROTATING EQPT SPARE PARTS				$618.0	$562.5		
PURCHASED EQUIPMENT COST				*6819*	*$14,237.0*	*$12,957.5*	*6818.98*	
SUBTOTAL (GENERAL FACILITY)				*19940*	*$43,498.6*	*$42,412.2*	*19940.30*	
SUBTOTAL (DILUTE ACID PRETREATMENT BLOCK)				*3083*	*$3,296.8*	*$4,017.2*	*3082.94*	
SUBTOTAL (ENZYME PRODUCTION)				*6205*	*$18,901.4*	*$8,592.6*	*6204.91*	
GRAND TOTALS (OVERALL FACILITY)				*29228*	*$65,696.8*	*$55,021.9*	*29228.15*	

Due to N Hinman
10-4-93

Interoffice Memorandum

National Renewable Energy Laboratory

TO:	Lisa Burns - 15/2	
FROM:	C. Wyman	Reviewer
THROUGH:	N. Hinman	Project Leader
DATE:		
SUBJECT:	Comments Back on Quality of Subcontract Deliverable	

Subcontract #:	11286-1
Subcontractor:	TDA Research
Subcontract Title:	Production of Ethanol from Lignocellulose by Separate Hydrolysis & Fermentaion: Case No. 5
Subcontract Deliverable:	Research Report
Summary Comments:	Excellent report. Fulfills need & document the process that introduces xylose fermentation to the wood ethanol technology.

(Please use separate sheet if more space is needed)

Detailed Comment Attached: Yes: No: X

Action Required: none

Action Required by Whom:

Action Required When:

File: BF/4.Subcontracts/ 11286-1.916 /memo & attachments

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98

NREL Research

September 10, 1993

Mr. Mark Yancey
National Renewable Energy Laboratory
1617 Cole Boulevard
Golden, Colorado 80401

Dear Mark:

Enclosed are two copies of the report: *Production of Ethanol from Lignocellulose by Simultaneous Saccharification and Fermentation: Case No. 5: Xylose Fermentation.* This report is the fifth deliverable on National Renewable Energy Laboratory (NREL) subcontract No. AA-2-11286-1.

Please distribute copies of the report to the interested technical staff at NREL. People who may have an interest include Paul Bergeron and Charlie Wyman.

Please call if you have any questions.

Sincerely,

John D. Wright
Vice President

JDW/meb

enc.

cc: Dan Feinberg
Project File

12345 West 52nd Avenue • Wheat Ridge, Colorado 80033 • (303) 422-7819 • FAX (303) 422-7763

NREL

Interoffice Memorandum

National Renewable Energy Laboratory

TO: Norm Hinman, Program Management Project Leader

FROM: Mark Yancey MY

DATE: September 16, 1993

SUBJECT: TDA Research Report: "Production of Ethanol from Lignocellulose by Simultaneous Saccharification and Fermentation Case No. 5: Xylose Fermentation" Subcontract No. AA-2-11286-1

Attached is a copy of the subject report from John Wright of TDA Research. Also attached are the standard transmittal forms for your use. Please review the report and return your comments to Lisa Burns by **October 14, 1993**. Thank you.

cc: L. Burns/subcontract file

Interoffice Memorandum

National Renewable Energy Laboratory

TO: ______

FROM: Norm Hinman

DATE: ______

SUBJECT: Review of Subcontractor Deliverable

Please review the attached subcontractor deliverable and comment back to me by __________ __________ as to the quality of the work.

Return your comments to me on the attached memo form after you sign as a reviewer. I will forward this memo form to Lisa Burns. You may keep the copy of the deliverable for your information.

Subcontractor #: 11286-1

Subcontractor: TDA Research

Title: Production of Ethanol from Lignocellulose by Separate Hydrolysis & Fermentaion: Case No. 5

Deliverable: Research Report

File: BF.4/Subcontracts/11286-1.916/memo only

b:\revsubdl.frm

Interoffice Memorandum

National Renewable Energy Laboratory

TO:	Lisa Burns - 15/2
FROM:	Reviewer
THROUGH:	Project Leader
DATE:	
SUBJECT:	Comments Back on Quality of Subcontract Deliverable

Subcontract #:	11286-1
Subcontractor:	TDA Research
Subcontract Title:	Production of Ethanol from Lignocellulose by Separate Hydrolysis & Fermentaion: Case No. 5
Subcontract Deliverable:	Research Report
Summary Comments:	

(Please use separate sheet if more space is needed)

Detailed Comment Attached: Yes: No:

Action Required:

Action Required by Whom:

Action Required When:

File: BF/4.Subcontracts/ 11286-1.916 /memo & attachments

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TDA Research

September 10, 1993

Mr. Mark Yancey
National Renewable Energy Laboratory
1617 Cole Boulevard
Golden, Colorado 80401

Dear Mark:

Enclosed are two copies of the report: *Production of Ethanol from Lignocellulose by Simultaneous Saccharification and Fermentation: Case No. 5: Xylose Fermentation.* This report is the fifth deliverable on National Renewable Energy Laboratory (NREL) subcontract No. AA-2-11286-1.

Please distribute copies of the report to the interested technical staff at NREL. People who may have an interest include Paul Bergeron and Charlie Wyman.

Please call if you have any questions.

Sincerely,

John D. Wright
Vice President

JDW/meb

enc.

cc: Dan Feinberg
Project File

12345 West 52nd Avenue • Wheat Ridge, Colorado 80033 • (303) 422-7819 • FAX (303) 422-7763